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Laboratory
Cincinnati OH 45268

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16. ABSTRACT

Treatment of wastewater from a subdivision in a physical-chemical treatment plant (screening, chemical coagulation, sedimentation, filtration, carbon adsorption, chlorination) was evaluated. The 190 m^3/day (50,000 gal/day) plant was housed in the shell of a standard house on a standard lot in a 127 home subidivision. During the 18 month evaluation period excellent treatment was achieved (99% removal of BOD_5, Suspended Solids, and Total Phosphorus). Shock loadings had almost no effect on plant performance because an equilization tank leveled out peaks and because of the ability of the physical-chemical processes to absorb excess loading. Extensive data on temporal characteristics of wastewater from a subdivision were collected during the evaluation. An experimental sludge filter and fluidized bed incinerator were installed to process the sludge but were not extensively used. The former did not function, the latter suffered from repeated mechanical breakdowns. Sludge was periodically hauled to a landfill by a septic-tank-hauler. Acceptance of the presence of a sewage treatment plant in the midst of the subdivision was excellent. No complaints of any type were registered by the homeowners. The system cost was higher than for a conventional plant. The actual construction cost exceeded $300,000 and operational expenses were greater than $0.53 per cubic meter ($2.00 per 1000 gallons). At the measured flow of 206 gal per home per day this represents a cost of $0.40 per home per day. It is anticipated that significant reductions in these costs would result from a redesign based on the experiences gained during the demonstration period.

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FOR
HOUSING AND COMMUNITY DEVELOPMENTS

by

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Levitt and Sons, Incorporated
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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the hazardous water pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

Provision of sewerage facilities for a dispersed population living in semi-isolated subdivisions is a significant problem. Either the subdivision must be connected to a central treatment plant, or a small scale plant must be provided for treatment of the community wastewater. The former procedure is often expensive because a long sewer run is required to serve only a few homes. Small scale biological treatment plants are also relatively expensive and exhibit poor performance when subjected to the wide fluctuations of loading inherent in a small flow situation. In this study a treatment plant for isolated subdivision wastes using wholly physical chemical treatment technology was evaluated. Physical chemical treatment systems are superior under fluctuatory loads, and are small enough to be placed in a shell of a typical suburban house, thus reducing land costs.

Francis T. Mayo, Director
Municipal Environmental Research
Laboratory

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ABBREVIATIONS

AC	-	Alternating Current
ALK	-	Alkalinity
BOD	-	Biochemical Oxygen Demand (Five-Day)
CFM	-	Cubic Feed Per Minute
COD	-	Chemical Oxygen Demand
Col	-	Coliform
DC	-	Direct Current
DIA.	-	Diameter
DO	-	Dissolved Oxygen
FTU	-	Formazin Turbidity Unit
GAL.	-	Gallon (United States Measure)
GPH	-	Gallons Per Hour
GPM	-	Gallons Per Minute
HP	-	Horsepower
Hr.	-	Hour
ID	-	Interval Diameter
JTU	-	Jackson Turbidity Unit (1FTU ≈ 1JTU)
mg/l	-	Milligrams Per Liter
MGD	-	Million Gallons Per Day
NH_3-N	-	Ammonia Nitrogen
P	-	Hydrolyzable Phosphorus
PSIG	-	Pounds Per Square Inch - Gauge
Res. Cl_2	-	Residual Chlorine
RPM	-	Revolutions Per Minute
SCFM	-	Standard Cubic Feed Per Minute
SS	-	Suspended Solids
TDH	-	Total Discharge Head
TDS	-	Total Dissolved Solids
TKN	-	Total Kjeldahl Nitrogen
TOD	-	Total Oxygen Demand
TS	-	Total Solids

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SECTION I

INTRODUCTION

GENERAL

The home and community building industry, in order to meet the housing requirements of today, has developed a concept of total community planning. Such community planning includes single-multi family housing, shopping centers, sites for industry, schools, churches, etc. Consequently the building industry is faced with a unique waste water disposal problem in that the industry, not the municipality, is generally required to provide adequate disposal facilities for new community developments. This trend toward total community planning, in which domestic and industrial wastes must be handled over a wide range of flow rates and waste characteristics, requires that building firms in this country have access to treatment facilities of types heretofore seldom utilized.

A national commitment to manage water quality has been manifested in the past decade in response to continual degradation of water supplies for beneficial uses. Technical, legislative, and financial moves have begun to zero in on the environmental problems. One of the first and most important needs in our national priorities to meet existing water management goals is to develop the minimum cost technology for removing contaminants from water and permanently disposing of them in order to alleviate water pollution problems. A further objective is to renovate waste waters for agricultural, industrial, recreational, or even potentially potable purposes. New advanced systems are now being designed to meet the new objectives of total pollution control, namely, the capability of treatment to extremely high levels of quality.

Advanced waste treatment has made great progress in supplying many tools to water resource management. Traditionally, small scale treatment plants have had to be installed where existing sewer lines were not available for hookup, or where barriers existed making hookup impractical. Suburban developments had to make use of so-called "package plants" to treat their waste water flows. These plants are usually adaptations of the conventional activated sludge process designed to meet specialized problems encountered in situations where flow is highly variable.

Two commonly used modifications are extended aeration and contact stabilization. Contact stabilization aerates waste water mixed with highly concentrated activated sludge to adsorb colloidal organics. Sludge from a clarifier is then aerated to stabilize the organic matter and renew the sludge surface for more adsorption. Extended aeration handles waste without primary settling. The aeration tank acts as an equalization basin to smooth out variations in load and to dilute slugs of concentrated or toxic impurities.

The most serious problems encountered with these package plant units are related to the highly varying natures of waste water flow together with waste strength or concentration. For example, it is common to see flow patterns where peak flows of 2.5 to 3 times the average occur during the day and zero flow occurs during hours of the night. Such flow and strength characteristics are ill suited to the capabilities of secondary biological processes. Coupling these shock loadings with substandard operation, maintenance, and reliability, it is not surprising that these plants cannot be expected to achieve, on a consistent basis, existing and projected water quality effluent requirements.

Physical-chemical treatment represents the first major innovation in sewage treatment in several decades. Such systems, while increasing the capability for handling a wide range of continuous and intermittent flow rates, offer added advantage in the ability to handle toxic industrial wastes and to remove components which cannot be treated biologically.

Advantages of a physical-chemical system include:

1. Smaller land area required (less than 1/2) of the conventional biological plant.
2. Higher degrees of treatment efficiency than conventional biological processes.
3. Lesser sensitivity to unusual loadings.
4. Lesser sensitivity to daily flow variations.
5. Greater design flexibility.
6. Greater operational flexibility and control.

Because these features make it possible to meet stringent effluent criteria in remote sites, physical-chemical treatment technology was chosen for a plant in a community in northern New Jersey. The basic concept was to place the treatment plant within the frame of a standard house on a standard lot in a subdivision. This procedure provides more land for ultimate development because a buffer zone need not be established around the treatment plant as it must if more conventional technology is used.

LOCATION

An advanced waste treatment system based on physical-chemical treatment technology was designed to serve approximately 127 single-family dwelling units, and a demonstration plant was constructed in a residential subdivision in Freehold Township, Monmouth County, known as Woodgate Farms. The nominal average design flow was 50,000 GPD.

PLANT DESCRIPTION

The plant design chosen provided for screening, flow equalization, chemical coagulation, sedimentation, filtration, carbon adsorbtion, and disinfection with chlorine. Sludge was to be dewatered on site and incinerated on site in a fluidized bed furnace.

OBJECTIVE

The specific objective of this project was to demonstrate the performance, economics, and applicability of a physical-chemical domestic wastewater treatment system designed to provide varying high quality discharges for isolated or developing communities having an average wastewater flow in the 25,000 to 500,000 GPD range.

APPROACH

The total project was conducted in the following three phases:

Phase I - Preconstruction Study and Engineering, during which all necessary preliminary studies, subsystem designs, and engineering plans were completed.

Phase II - Construction, including purchase of all equipment, subcontract work, construction of the building, and assembly of the complete system.

Phase III - Operation and Evaluation, including a start-up and break-in period while houses in the development were being built and occupied (October 1972 through June 1973), and a period of operation under steady state conditions (July 1973 through March 1974) to obtain data on all aspects of plant performance.

SECTION II

CONCLUSIONS

1) The physical-chemical wastewater treatment system provided a high-quality discharge while processing all the sewage from an isolated community of 127 single-family dwelling units over a period of 18 months.

2) Average treatment performance showed removals of 99% BOD$_5$, suspended solids and phosphorus; 37% Kjeldahl nitrogen. The use of the ferric chloride coagulant caused reduction of 46% alkalinity and a threefold increase in chlorides.

3) Unit operations performance shows:

a. Primary screen removed approximately 1/4 of the BOD, suspended solids and phosphorus.

b. Chemical treatment and clarification removed approximately 2/3 of the BOD$_5$, suspended solids and phosphorus.

c. Carbon adsorption removed about 16% of the BOD and 25% of the TKN.

d. Filtration was relatively ineffective.

4) Characteristics of raw sewage from the totally domestic source were:

Flow - 206 gpd/unit
BOD - 207 mg/l
SS - 242 mg/l
TS - 628 mg/l
P - 10.8 mg/l
TKN - 44.7 mg/l
Coliforn - 1.1×10^6 MPN/100 ml

5) Twenty-four hour raw sewage profiles on flow and eight analytical parameters showed substantial variation. For example: Hourly peak flow (180% of average) occurred at 11 a.m.-noon, with secondary peaks at 8 p.m. and 11:30 p.m. Minimum flow was registered between 4 and 5 a.m. at 14% of average. Analytical parameters showed similar variation but with peak and minimum concentrations occurring at different hours.

6) The surge tank, provided to level out raw sewage variation, was very effective. It allowed constant flow rate processing through the plant with only periodic adjustment. Also variations in the concentration of major constituents were smoothed out and reduced to about half that in the instantaneous flow.

7) Chemical consumptions were higher than anticipated, averaging 267 mg/l for $FeCl_3$; 154 mg/l for NaOH; 2.6 mg/l for polyelectrolyte; and 26.4 mg/l for chlorine. They have been reduced with plant modifications which include more automation, pH control of $FeCl_3$, and greater mixing effectiveness and detention times.

8) Operating costs were higher than anticipated because:

a. Labor costs were excessive due to the experimental nature of much of the equipment, due to the high level of analytical work carried out, and because more manual control was required than anticipated.

b. Chemical, fuel and utility costs escalated significantly during the trial period.

9) Solids disposal.

a. Sludge discharged from the clarifier averaged 6.8% solids.

b. Sludge discharged from the primary screen averaged 17% solids.

c. An experimental continuously regenerating filter did not perform satisfactorily for dewatering of the sludges.

d. Direct incineration of clarifier sludge was successfully conducted for short periods, but required significant quantities of fuel.

e. Particulate emission was 0.011 grains/cu. ft. from the incineration.

f. The incinerator was plagued with problems such as failure of the distributor plate and hot spots in the shell. Its use was discontinued due to above problems coupled with high fuel costs.

10) An automated sludge removal system using an optical density probe has worked exceptionally well, and is a recommended feature for future design considerations.

11) The magnetic filter used here has limited potential for use if applied after a clarification process which functions properly.

12) The turbidimeter used to signal for recycle of poor quality effluent to the head of the plant during periods of non-operator attendance proved highly satisfactory. If possible, however, the turbidimeter should have a continuously adjustable set point in order to be used to full advantage by the operator.

13) The plant had an attractive appearance with no odors or problems with close neighboring houses in four years of operation.

14) Liquid handling, chemical feed, mixing, and control thereof, can be highly automated to minimize manpower requirements.

15) Disadvantages of this type of treatment plant are:

a. High electric and energy cost
b. Sludge disposal problem
c. Deficiency in nitrogen removal
d. Pumps and equipment are more prone to maintenance problems than in large plants.
e. High cost of chemicals.

SECTION III

RECOMMENDATIONS

Based on the experience gained from research, development, construction and evaluation of this physical-chemical waste water treatment plant, several positive steps are recommended to increase plant efficiency and reduce operating costs.

Plant Efficiency: Positive head pumps in the lift station would provide more dependable operation; solids cutting capability in raw sewage pumps would improve settling process in the clarifier; air mixing provisions in surge tank would produce a better mixture and a larger surge tank storage capacity would enhance equalization of BOD loading.

Operational Cost Reduction: Manpower requirement can be minimized by highly automating liquid handling, chemical feed, and mixing. Bulk storage provisions for chemicals would reduce chemical costs significantly. In addition, spray washing the screens with plant effluent would minimize water consumption costs.

Within the carbon column, biological regeneration of carbon seemed to be evident. Investigation into this phenomenon could result in development of a system which does not require thermal regeneration of activated carbon. It is also recommended that the volume of the carbon quench tank and the carbon spent tank be enlarged to the equivalent volume of carbon within the column. These tanks could then serve as carbon storage tanks when the column must be emptied.

Carbon columns are in need of sophisticated design considerations not previously highlighted in the literature. Among these are:

a. Adequacy of backwash
b. Proper distribution of support mechanisms
c. Proper design of take-off
d. Properly designed screens (stainless steel in screens and support mechanisms may have limited life).

Sludge disposal methods specifically applicable to plants of this type and size range are needed:

a. Land fill facilities in the nearby vicinity is the most logical and probably economical solution.

b. A large treatment plant nearby normally handling scavenger and industrial wastes would be a good alternative.

c. Dewatering and land disposal is a current possibility but has an uncertain future.

d. Dewatering and incineration is not economical unless the plant is in the 500,000 gpd range.

Any dewatering system must recognize the difficulties of accomplishing this on ferric sludges. The system tried at Freehold consisting of the coil vacuum filter had limited success probably due to lack of adequate pre-treatment and high chemical levels at the time of experimentation.

The concept of putting a treatment facility in the shell of a home in a subdivision should be tried again with a different design in another location. Special attention should be given to the physical arrangement in the shell, the need to use equipment and construction features different from those in standard home construction, and the method of sludge disposal. Treatment processes which have not been proven at the scale to be used should not be included. Special attention should be given to the problems which may result from the use of small site lines and small capacity pumps.

SECTION IV

PHASE 1. PRECONSTRUCTION STUDY AND ENGINEERING DESIGN

A. LIQUID HANDLING PROCESS

Predesign studies on all liquid process unit operation were conducted in a transportable pilot plant, which for these studies was installed at the Township of West Goshen, Pa., Sewage Treatment Plant. The West Goshen treatment plant is a nominal 3 MGD trickling filter plant, treating sewage from the bedroom community of West Goshen along with a small percentage of commercial wastes.

The sewage entering this plant is characterized as weak, with BOD and suspended solids averaging about 100 mg/l. Infiltration of ground water into the collection system is estimated to be 50-60% of the total flow. The typical diurnal flow pattern entering the plant is shown in Figure 1.

The transportable pilot plant contained the following unit process equipment:

1. Primary Screen
2. Surge Tank
3. Chemical Addition System
4. Magnetic Filter
5. Carbon Adsorption System

Figure 2 is a schematic drawing of the arrangement of these components in the process excluding the carbon adsorption system which followed the magnetic filter.

The primary screen was a Bauer Hydrasieve. It was operated with a number of different screen openings ranging from 10 to 40 mils. A 20 mil opening was found to be optimum to give acceptable solids removal with a minimum of blinding.

The liquid from the screen flowed by gravity into the surge tank which was a portable 4,000 gallon vinyl-lined swimming pool.

The screened sewage was pumped through a flow control valve and flow meter into the treatment system where inorganic flocculant was added continuously. Previous work had established ferric chloride as the most cost effective inorganic flocculant. Kenics

FIGURE I

WEST GOSHEN, PA.
WASTE TREATMENT PLANT
FLOW DATA
FOR FIFTY DAY PERIOD

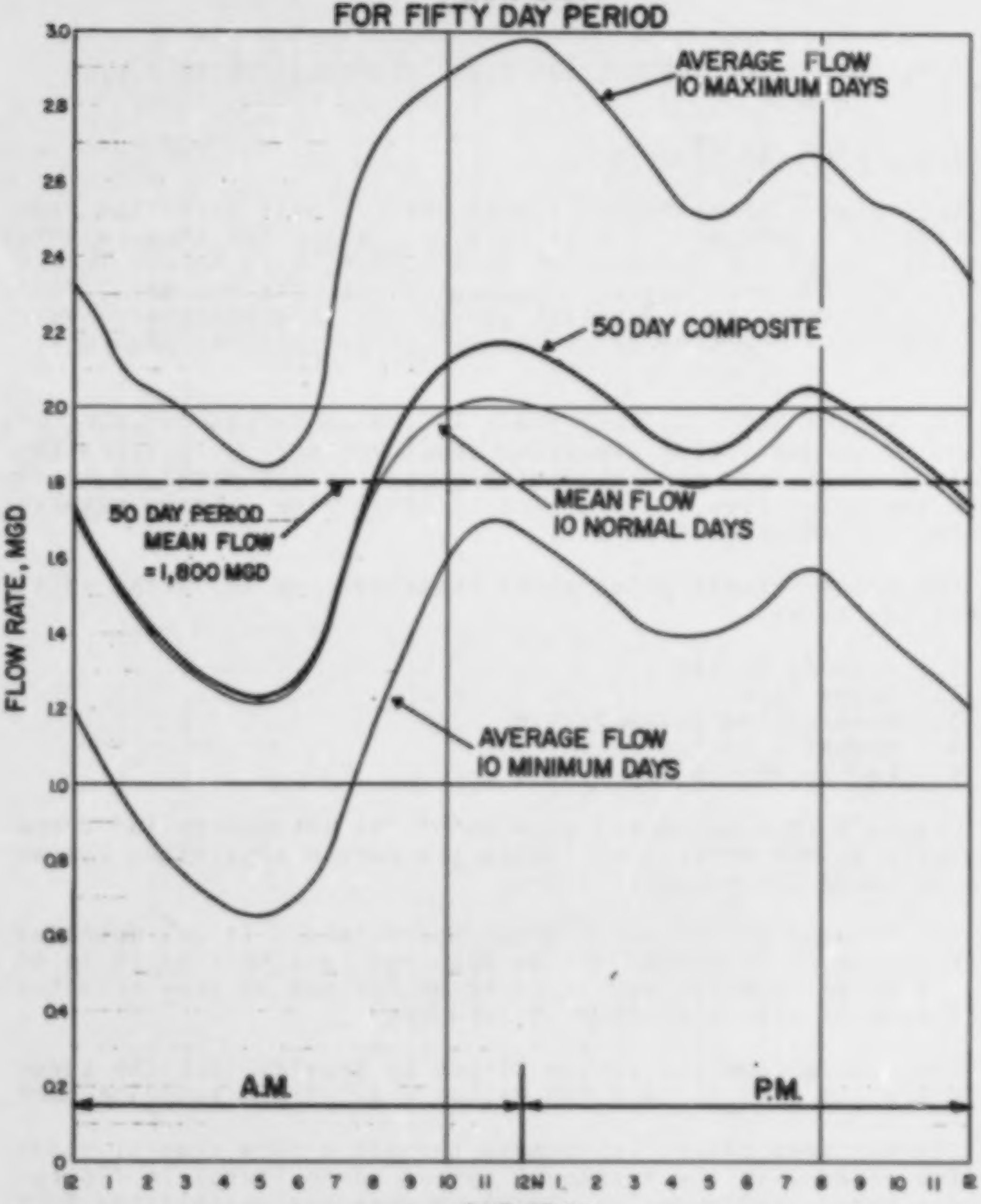
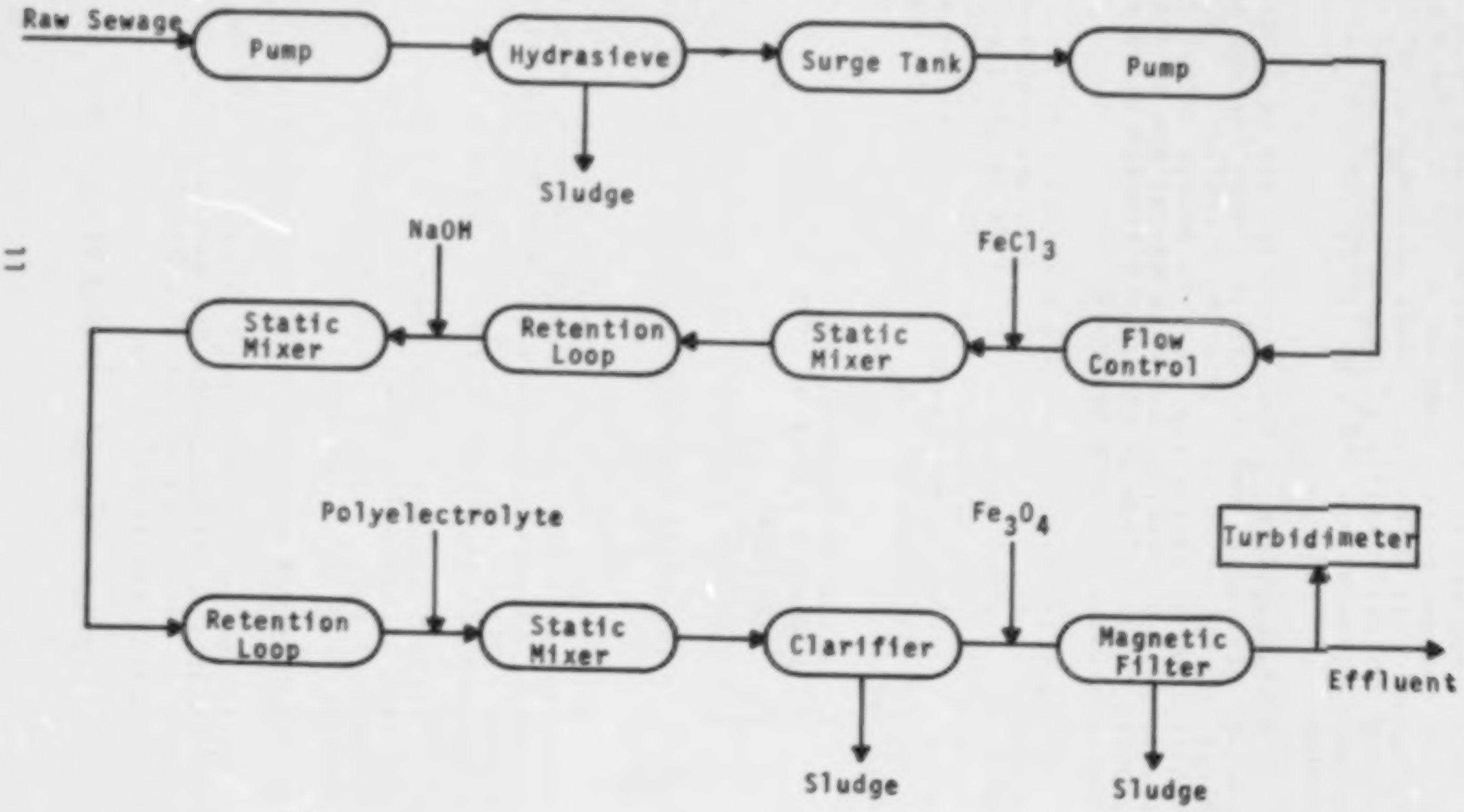


FIGURE 2

WEST GOSHEN PILOT PLANT OPERATION
SCHEMATIC FLOW DIAGRAM


Raw Sewage
Pump
Hydrasieve
Sludge
Surge Tank
Pump
Flow Control
$FeCl_3$
Static Mixer
Retention Loop
NaOH
Static Mixer
Retention Loop
Polyelectrolyte
Static Mixer
Clarifier
Sludge
Fe_3O_4
Magnetic Filter
Sludge
Turbidimeter
Effluent

Corporation Static Mixers were used to achieve satisfactory mixing with minimum energy, followed by a variable retention loop to attain optimum contact time for ferric phosphate precipitation and hydroxide floc formation. Study of flocculation process established the desirability of controlling the pH of the stream after primary coagulation, and a pH control section was added to the system.

Following primary floc formation and pH control a polyelectrolyte was added, and the stream flowed through another Static Mixer and retention loop into the clarifier. The clarifier was constructed of plexiglass sheet to enable visual observation of the effects of chemical and flow variations on the clarification process. The action of the sludge blanket under varying operating conditions was also observed.

The filtration device was a magnetic filter, "Ferrofilter", supplied by the S. G. Frantz Co. Its use required the addition of finely divided magnetic iron oxide to the clarified stream. Contact with the magnetic field in the filter removed these iron oxide particles as well as occluded residual suspended solids from the clarification step. Several variables were studied with this device:

1. Field Strength
2. Amount and type of iron oxide
3. Throughput
4. Filter loading
5. Backwash mode and cycle time

Effluent turbidity from the magnetic filter was continuously recorded.

Following filtration a portion of the clarified stream was contacted with granular activated carbon in 2-inch ID adsorption columns. An upflow expanded bed mode of operation was utilized. Carbon capacity, shape of the column exhaustion curve, and optimum contact time were studied in these columns.

Table 1 summarizes the range of operating conditions used in the West Goshen pilot plant to obtain design information for the Freehold plant. Table 2 presents a summary of typical operating results from the pilot plant.

B. DESIGN PARAMETERS AND EQUIPMENT SELECTION

Following the pre-design research at the West Goshen plant, the following design parameters were selected for the plant at Freehold, based on anticipated population, State regulations and pilot plant analytical data.

1. Design Capacity - 50,000 GPD (35 gpm)

TABLE 1

WEST GOSHEN PILOT PLANT OPERATION OPERATING CONDITIONS

	Range	Typical
Flow Rate from Surge Tank, gpm	10-27	20
Fe^{III} Addition, mg/l	19-71	45
Polyelectrolyte Addition, mg/l	0.75-2.0	1.0
Fe_3O_4 Addition, mg/l	5-28	10
Carbon Column Feed Rate, $gpm/ft.^2$	3-9	5
Carbon Contact Time, min.	20-60	40
Hydrasieve Screen Opening, mils	10 to 40	20

TABLE 2

WEST GOSHEN PILOT PLANT OPERATION

OPERATING RESULTS

Analysis, mg/l	Suspended Solids	Soluble Ortho-Phosphate	Total Organic Carbon	Biochemical Oxygen Demand
Surge Tank	98	18. 5	44	95
Clarifier Effluent	14. 9	1. 0	18. 1	-
Magnetic Filter Effluent	7. 5	0. 8	16. 3	23. 4
Carbon Column Effluent	3. 2	0. 5	1. 5	4. 6
		PER CENT REMOVAL		
Clarifier	85	94	59	-
Filter	92	96	63	75
Carbon Columns	97	97	97	95

2. Lift Station

Dual pump system with automatic level controls. A vacuum-primed unit was selected with dual centrifugal pumps, each with capacity of 75 gpm.

3. Primary Screen

Bauer Hydrasieve - 36" wide x .020" openings. Since it was possible for both lift station pumps to operate simultaneously, the Hydrasieve was sized for a flow of 150 gpm. In order to avoid corrosion problems, the screen was constructed of 316 stainless steel and the frame of fiberglass reinforced polyester.

4. Hydrasieve Enclosure

Because of the possibility of generating odors around the Hydrasieve and the necessity for frequent cleaning in the area it was decided to enclose this unit. A room was built with Transite paneling around the Hydrasieve on three sides and a 6 ft. glass sliding door on one side. This enclosed area was ventilated by a small exhaust fan (1/8 HP, 195 CFM) discharging to the atmosphere.

5. Surge Tank

15,680 gallons, reinforced concrete, 13.5' x 13.5' x 11.5' deep. The design capacity of the surge tank was 11,500 gallons. This was based on operation at constant withdrawal of 36 gpm with the flow profile as shown by Figure 3. This profile was derived from flow data of a Levitt and Sons' Stony Brook, Long Island, New York, treatment plant processing domestic sewage which contained essentially no infiltration. Under these conditions maximum calculated accumulation is 11,500 gallons. Thus, the surge tank has excess capacity of 4,180 gallons or 36% of normal requirements.

6. Surge Tank Agitator

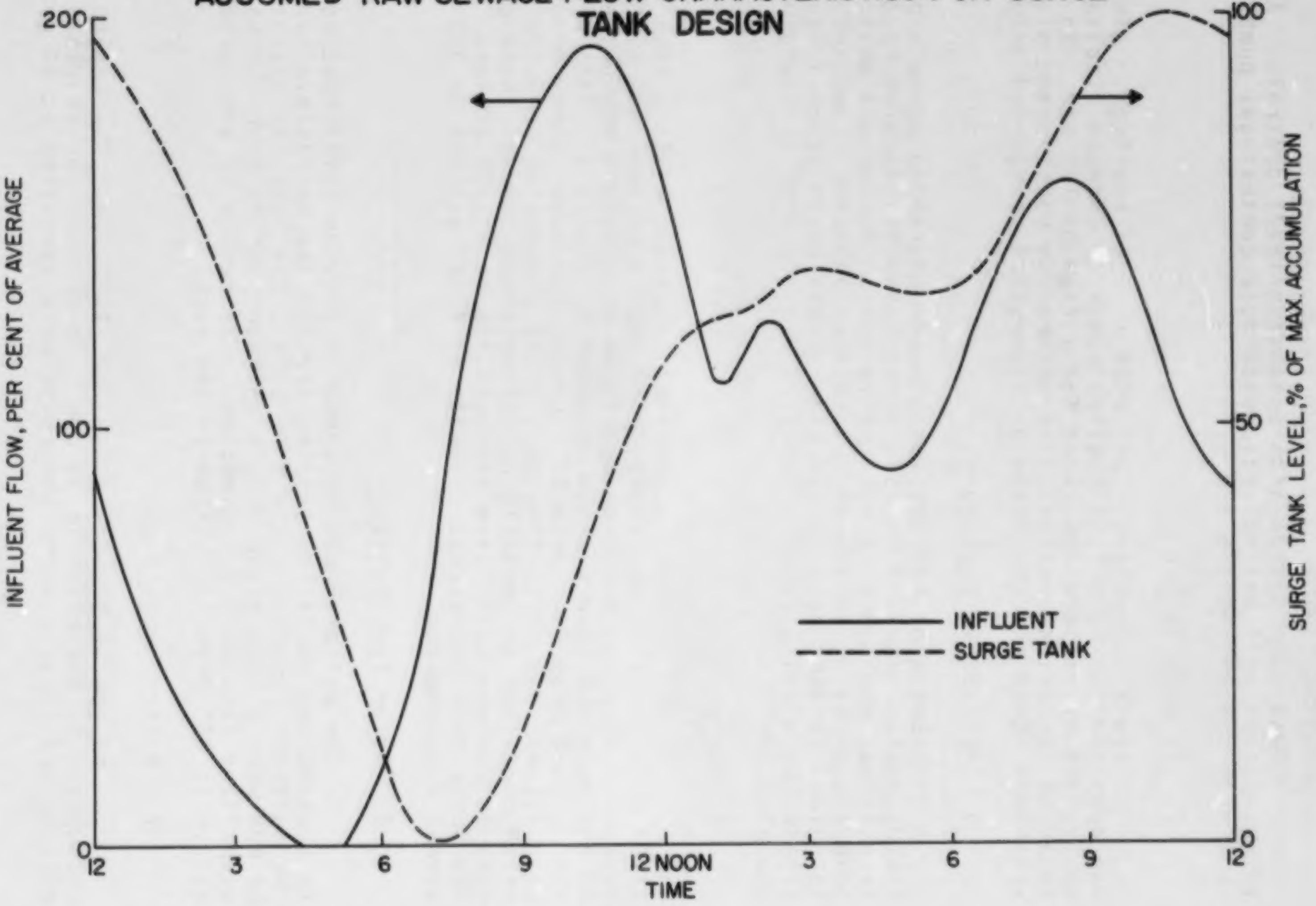
The agitator was selected to provide sufficient agitation to suspend any particles passing through the Hydrasieve screen. The agitator selected was a 2 HP unit. The impeller operates at 68 RPM with a four blade single agitator 36" in diameter with stabilizer fins on a 24" diameter. The center of the agitator paddle is 24" from the bottom of the tank.

7. Primary Feed Pumps

50 gpm at 120 ft. TDH, 3 HP, 2450 rpm. The design throughput of the plant is 35 gpm. To provide for overload conditions and recycle water the pumps were specified at 50 gpm.

FIGURE 3

ASSUMED RAW SEWAGE FLOW CHARACTERISTICS FOR SURGE TANK DESIGN



Although the head requirement for the primary feed pumps was approximately 50 TDH, the pumps were specified to have a 120 ft. TDH so that they would be interchangeable with the adsorber feed pumps.

8. Chemical Treatment System

Chemical preparation and pumping equipment, mixers, and retention loop. This system included dual 50 gallon polyethylene tanks for $FeCl_3$ and polyelectrolyte and one tank each for NaOH and magnetic iron oxide. Each tank was provided with an agitator. Chemical feed pumps were electronic drive diaphragm pumps by Precision Chemical Pump Company. Electronic controllers were provided for the $FeCl_3$ and NaOH pumps. Kenics Corporation Static Mixers, PVC-2" pipe size by 8 elements, were installed after each chemical injection point. The retention loop was designed for one minute detention at design flow, and consisted of 55 feet of 4" PVC pipe mounted horizontally.

Calculation of loop length:

1 min. x 35 gallons/minute = 35 gallons = 4.68 ft.^{3}
Retention pipe loop at 0.085 ft.^{2} for 4 in.
PVC pipe = 4.68 ft.^{3} $\div$ 0.085 ft.^{2} = 55 ft. long

9. Chemical Requirements

Based on raw sewage composition at design flow: based on selected design criteria agreed upon by all agencies in the approval process

	mg/l
BOD_5	275
Suspended Solids	275
Phosphorus	12

$FeCl_3$: two mole ratio to phosphorus

$2 \times 12 \times 162/31 = 125$ mg/l $FeCl_3$ = 48 lbs./day

NaOH: to neutralize half of $FeCl_3$

$FeCl_3 + 3\ NaOh \rightarrow 3\ NaCl + Fe\ (OH)_3$

$125/2 \times 3(40)/162 = 46$ mg/l NaOH = 19 lbs./day

Hercofloc: 1 to 2 mg/l

1 to $2 \times 10^{-6} \times 0.050 \times 10^6 \times 8.35 = 0.42$ to 0.83 lbs./day

Chlorine: 10 mg/l

$10 \times 10^{-6} \times 0.050 \times 10^6 \times 8.35 = 4.2$ lbs./day

Fe_3O_4: 10 mg/l

$$10 \times 10^{-6} \times 0.050 \times 10^6 \times 8.35 = 4.2 \text{ lbs./day}$$

10. Clarifier

1.5 gpm/ft.^{2} upflow rate and 30 minutes retention time at design flow rate. Actual fabrication:

6' dia. x 6' side-wall depth with 1'6" dia. center well
Upflow area = 28.3 ft.^{2} - 1.8 ft.^{2} = 26.5 ft.^{2}
Volume = 1270 gallons
Upflow Rate = 1.32 gpm/ft.^{2}
Retention Time = 36 minutes

Several commercially available clarifiers were investigated for utilization in the Freehold plant. The size required was below the normal size of most clarifier manufacturers and consequently the cost of commercial units was very high. It was then decided that it was more economical to design a special clarifier for this plant and have it fabricated. As originally designed the clarifier had a conical bottom sloping down to the center line. Equipment arrangement studies showed, however, that there was not enough elevation to locate the clarifier above the control panel and still have access room to the rake drive. The slope of the bottom was, therefore, changed to the outside instead of to the center line. This change allowed the control panel equipment to fit beneath the clarifier. Because of the limited ability to lift tanks in the building the clarifier was made in two sections and the center seam welded in the field. The main tank was fabricated from steel. The inlet well and overflow weir were constructed of 304 stainless steel.

11. Clarifier Rake

Tip speed: 1.1 in./sec. The clarifier rake was driven by a gear reducer with an output of 0.67 rpm. The rake was constructed of steel pipe and angle iron. Because of the limited headroom, it was necessary to use a gear drive with as low a profile as possible.

12. Sludge Level Controller

Ultrasonic sensors with adjustable depth. During the development stages at the West Goshen pilot plant the sludge level was normally controlled manually. An ultrasonic level detector was tested for use on the sludge level control mechanism, and this equipment was installed at Freehold.

13. Sludge Pumps

Jabsco positive displacement pumps had been used successfully at West Goshen for handling sludge from the clarifier. These same pumps were utilized in the design of the Freehold plant.

14. Magnetic Filter

Upflow rate 50 gpm/ft.^{2} at design flow. Flush Cycle: Adjustable with timer control. Typical operating cycle: 1 hr. interval; 20 sec. flush; 20 sec. rinse. A Model 473 Ferrofilter (S.G. Frantz Company) with 12" ID was used for this application.

15. Adsorber Feed Tank

7 ft. dia. x 7 ft. deep; operating capacity 1800 gal. This tank was constructed of carbon steel.

16. Adsorber Feed Pumps

Identical to Item 7.

17. Carbon Adsorption System

After considering various proposals by potential suppliers, it was decided to purchase a carbon adsorption system from Chemec Process Systems Inc. This unit consisted of the carbon adsorber, the quench tank, the blow case, interconnecting piping, pneumatically operated control valves, and a panel for activating the control valves. In order to avoid corrosion from the activated carbon, all piping was made of type 304 stainless steel.

18. Activated Carbon Adsorption Column

Empty bed contact time: 40 minutes; Upflow rate: 2.2 gpm/ft.^{2}; Actual fabrication: 4.5 ft. dia. x 14 ft. straight side x 90° top cone x 60° bottom cone. Total volume 253 ft.^{3} = 1890 gallons; contact time = 1890/35 = 54 minutes.

The original adsorption system concept consisted of three downflow pressure columns in series. A study of other systems, coupled with theoretical considerations and physical constraints, indicated that a single upflow packed-bed column using pulse transfer would have several advantages. These included more efficient overall operation, substantially less space requirements, a simpler piping system, and significantly lower total cost. Off-setting these advantages were concern about suspended solids penetration into the column, the risks involved with only a single contact unit, and difficulties experienced with upflow packed-bed operation at other locations.

A compromise solution was reached retaining the single column concept but increasing the straight side of the column from 12 to 14 feet to allow operation in an upflow expanded-bed mode if desired.

The column was constructed of steel, lined with Plasite No. 7155 with a minimum dry thickness of 8 to 10 mils. The bottom cone included a Neva-Clog underdrain constructed of stainless steel. Design pressure was 50 psig.

19. Spent Carbon Blow Case

30 ft.^{3} to contain approximately 5% of carbon column charge per transfer (assuming blow case to be half full of carbon). Actual fabrication: 3'-6" dia. x 3' straight side, dished heads; Total volume-36 ft.^{3}. The blow case was constructed of steel with Plasite No. 7155 lining. Design pressure was 50 psig.

20. Carbon Quench Tank

5 ft. dia. x 5 ft. x 60° cone bottom; volume-930 gallons. The quench tank was constructed of type 304 stainless steel with an internal launder at the top to allow backwashing of the carbon.

21. Spent Carbon Tank

4.5 ft. dia. x 6 ft. x 60° cone bottom; volume-850 gallons. The spent tank was built of carbon steel and lined in the field with two coats of Ceilcote Flakeline No. 252 and a top coat of clear resin. It was also equipped with an internal launder at the top to allow for excess water overflow and for backwashing of the carbon.

22. Sludge Holding Tank

Capacity for approximately 2 days sludge storage at 3% solids. 5 ft. dia. x 8 ft.=1150 gallons. The sludge hold tank was designed to have a working capacity of 1100 gallons and constructed of carbon steel. The original concept was to sluice the sludge out of the bottom of the sludge hold tank. A special bottom was therefore designed with a water inlet and sludge outlet nozzle 180° apart. This concept was later abandoned but the tank specifications had already been prepared and since the additional nozzle was not detrimental, it was left in.

23. Chlorine Feeder

In order to avoid having chlorine under pressure in the building, a Wallace and Tiernan chlorine injection system was used. The driving force for the injection unit was a 10 gpm rotary pump. The chlorine cylinders, chlorine scales, and feed rotometers were located in a separate room in the building as required by New Jersey law. This room was located behind the laboratory and office space so that the operator could observe the consumption of chlorine, as shown on the scales, through a window.

24. Chlorine Contact Tank

30 minutes at design rate =30x35= 1050 gallons. Actual Fabrication: 5 ft. dia. x 9.5 ft.=1200 gallons. In order to encourage plug flow the tank was baffled, with flow entering the top of the tank on one side of the baffle, then down under the baffle, up, and out the overflow nozzle. It was constructed of carbon steel. There was some concern about overchlorinating in case the flow stopped. Consequently, a weir was put in the overflow nozzle so that when the flow stopped the level in the weir would drop. This in turn would close a switch which would shut off the circulating water pump and thus stop the input of chlorine into the chlorine contact tank.

25. Instrumentation

Economical operation of the plant required that operator attention be at a minimum. Consequently, the extensive intrumentation supplied was approximately 12% of capital cost. The major instrument systems for the liquid handling process were as follows:

a. Level alarm showing high level in the wet well
b. Level recorder alarm on the surge tank
c. Process flow recorder controller alarm
d. pH recorder controller alarm for NaOH addition ahead of the clarifier.
e. Level alarm showing high level in the sump
f. Level controller adding make-up water to the adsorber feed tank.
g. Turbidity recorder controller alarm, setting the ferric chloride flow as determined by the turbidity leaving the clarifier, and also diverting flow from the clarifier to the surge tank in case of high turbidity.
h. Timers on the magnetic filter to set the time and frequency of backwash.
i. Effluent flow recorder and totalizer.
j. Sludge level controller in the clarifier to automatically control discharge of sludge to the holding tank.
k. Telephone connection to transmit any alarm condition to a remote location.

B. SOLIDS DISPOSAL

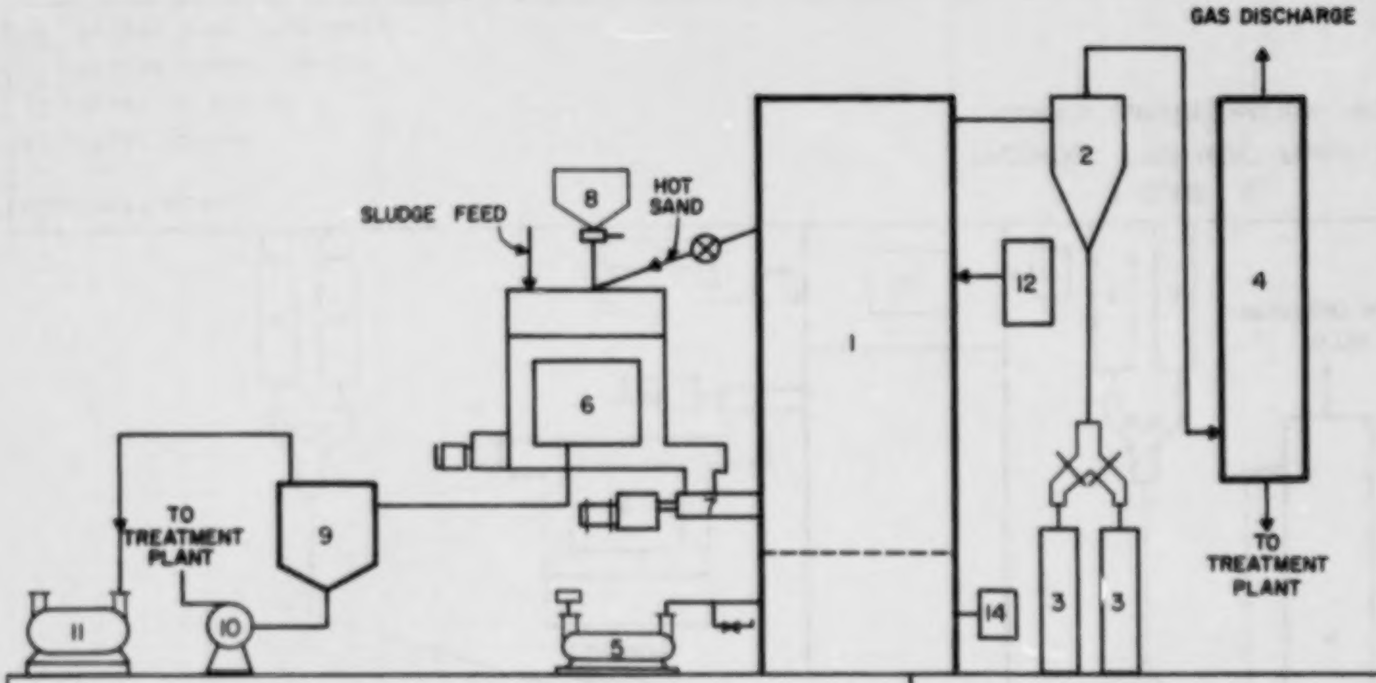
The solids disposal segment of the process was in a considerably less advanced state than the liquid handling process. It was based upon concepts developed by Procedyne Corporation, New Brunswick, New Jersey. This concept was based upon the use of a fluidized bed reactor in both incineration and carbon regeneration modes. In addition, the raw sludge would be mixed with sand from the fluid bed reactor and the mixture dewatered in a continuous sludge dewatering filter of novel design. After dewatering the mixture of sand and sludge would be fed to the incinerator. The combustible portion of the sludge would be burned off and the sand recycled to the filter. This process is shown schematically in Figure 4. The Carbon Regeneration mode of operation is shown schematically in Figure 5.

The sludge dewatering concept shown in Figure 4 consisted of continuously pumping sludge from the Sludge Holding Tank into the top of a filter (6) where the sludge would mix with hot sand recycling from the Incinerator. Make-up sand could be added as required from a bin (8). The sand-sludge mixture would move down the filter screen where the liquid would be removed under vacuum to a filtrate receiver (9) by a vacuum pump (11) and returned to the head of the plant by a transfer pump (10). The dewatered sand-sludge mixture would be discharged to a feed screw (7) which would convey it into the fluid bed of the Incinerator (1). Fluidizing and combustion air would be supplied by a blower (5). Fuel oil would be injected automatically via gun burners as required to hold the desired temperature of 1450°F. (12). Combustion gases and particulate matter from the Incinerator would pass through a dry cyclone (2), where the ash could be separated and collected in bins (3), then to a scrubber (4). The clean gases would be discharged to the atmosphere and the scrubber water returned to the head of the plant.

It was contemplated that once a month sludge burning would be interrupted and the Incinerator used to regenerate spent carbon from the carbon absorber. In this concept, shown in Figure 5, the sand filter would be run empty, sand recycle stopped, and the gun burners shut down. A plenum burner system (15) would be used to supply heat and fluidizing gases which would thermally regenerate the carbon yet would be too low in oxygen content to oxidize the carbon. The spent carbon from the Spent Carbon Storage Tank would be dropped through the filter to the feed screw (7) and fed into the fluidized bed of sand. Temperature would be controlled at 1800°F. by the plenum burner system (15). Regenerated carbon would be stored under water until needed to refill the Carbon Absorber. The Incinerator off gases would exit via a dry cyclone (2) as before then through an oil fired after burner (13) to oxidize any hydrocarbons. The gases would be scrubbed to remove particulates then discharged to the atmosphere.

1. Continuously Regenerating Filter

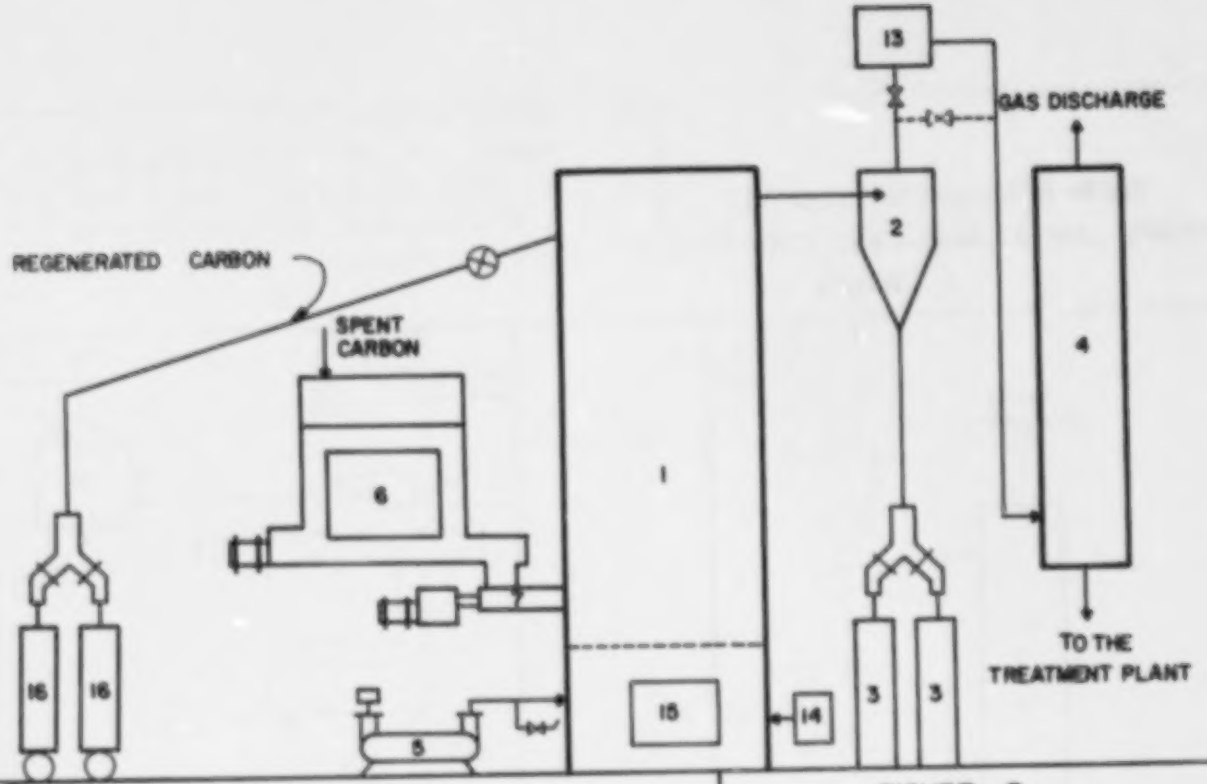
The development of the continuously regenerating sand filter proved to be the most technically difficult part of the program. The filter concept was experimental and based upon the concurrent utilization of three different dewatering techniques: evaporation, vacuum filtration, and compression. Preliminary work had been done by Procedyne on the dewatering of Paunch Manure, for the Earl Morall Co. an Illinois meat packer, using this technique. Continuation of this program using primary clarifier sludge for dewatering studies led to the following conclusions.


GAS DISCHARGE
SLUDGE FEED
HOT SAND
TO TREATMENT PLANT
TO TREATMENT PLANT
1
2
3
3
4
5
6
7
8
9
10
11
12
14

EQUIPMENT LIST:

1. Incinerator
2. Dry Cyclone
3. Ash Bins
4. Scrubber
5. Blower
6. Sand Filter
7. Feed Screw
8. Sand Feed Port
9. Filtrate Receiver
10. Transfer Pump
11. Vacuum Pump
12. Fuel Oil System
14. Start-Up Burner

FIGURE 4
FREEHOLD TREATMENT PLANT FURNACE
SLUDGE INCINERATION MODE


13
GAS DISCHARGE
2
4
REGENERATED CARBON
SPENT CARBON
6
1
16
16
15
14
3
3
5
TO THE TREATMENT PLANT

EQUIPMENT LIST:

13. After Burner
14. Start-Up Burner
15. Plenum Burner System
16. Carbon Bins (Mobile)

FIGURE 5
FREEHOLD TREATMENT PLANT FURNACE
CARBON REGENERATION MODE

a. Vacuum is a more effective driving force than external pressure in the filtration of these slurries.

b. Filtration rates average between 0.2 gpm/ft.^{2} and 0.6 gpm/ft.^{2}.

c. The filtration device should be constructed to provide a contact and mixing section, a filtration section, and a cake removal section.

d. The filtration section should be truncated in shape for smooth and continuous downward motion of the slurry. The angle of truncation should be 65°-75° to prevent bridging of the wet sand/sludge mixture.

e. A multi-screw live bottom should be provided for dry cake removal.

These considerations in combination with material and heat balance calculations involved in the dewatering process led to the design and specifications shown in Figure 6 and as described herein. In this design, evaporation occurs in the mixing zone above the distribution screw. Filtration takes place through the inner sides of the unit. This false side is constructed of stainless steel wedgewire screen with 10 mil openings. Compression dewatering occurs beneath the three feed screws through a false bottom screw which is also constructed of 10 mil wedgewire.

Initial demonstration tests with this concept showed that the volume of dewatered sand-sludge mixture was too small for controllable discharge with the three-screw live-bottom-bin configuration. This led to a further refinement in the design. The three removal screws were replaced by a single screw placed in a drop section in the bottom of the unit. This was fed by two counter rotating paddles to prevent bridging of the wet sand. This modification is shown in Figures 7 and 8. The unit was built in this manner. Prior to installation of the filter at the Freehold Plant, the unit was subjected to intensive testing. Several mechanical and operational problems were encountered and corrected:

a. Sand Sludge Distribution

Tests on the filter as originally delivered showed that sand-sludge distribution into the top of the filter was not uniform and resulted in piling up of the sand toward one end of the filter. This problem was resolved by replacing the top ribbon distributor with a variable pitch paddle arrangement. The lower paddles were also replaced with an adjustable unit so that the proper arrangement could be determined experimentally.

b. Sand Leakage

In the original experiments sand leaked past the

PARTS IDENTIFICATION:

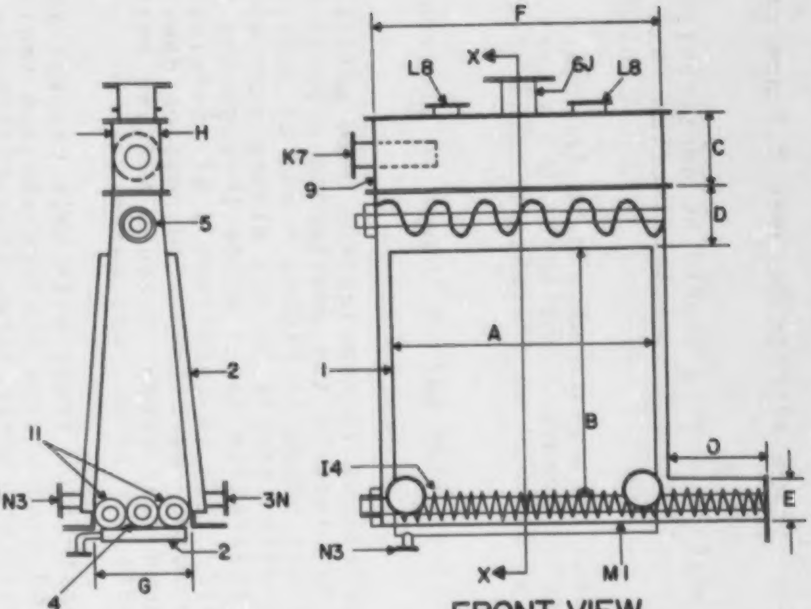
1. Screens
2. Filtrate Collector
3. Filtrate Outlet Nozzle
4. Short Pitch Screw (Feed Screw)
5. Distribution Screw
6. Sand Inlet Nozzle
7. Sludge Inlet Nozzle
8. Vent & Insp. Nozzles
9. Sludge Mixing Zone
10. Paddle Screws

DIMENSIONS:

A Screen Width------------------18"
B Screen Height-----------------12"
C Mixing Zone Height------------ 9"
D Bed Leveling Zone------------- 6"
E Trough Diameter--------------- 4"
F Filter Width------------------21"
G Filter Depth------------------13"(max)
H Mixing Zone Depth------------- 4"
I Short Pitch Screw Size-------- 3"
J Sand Inlet Nozzle Dia.--------- 3"
K Sludge Inlet Nozzle Dia.------- 1"
L Vent Size--------------------- 2"
M Bottom Screen Size------------18"x12"
N Filtrate Outlet Nozzle Dia----- 2"
O Feed Length-------------------To Be Specified

AREAS:

Total Filtration Area Provided - 5 Sq.Ft.



SECTION X-X

FRONT VIEW

PROCEDYNE CORPORATION
NEW BRUNSWICK, NEW JERSEY
JOB No. 1431 - FIGURE 6
FREEHOLD TREATMENT PLANT
REGENERATING SLUDGE
DEWATERING SAND FILTER
DATE 11-2-71
DRAWING No. B-05552

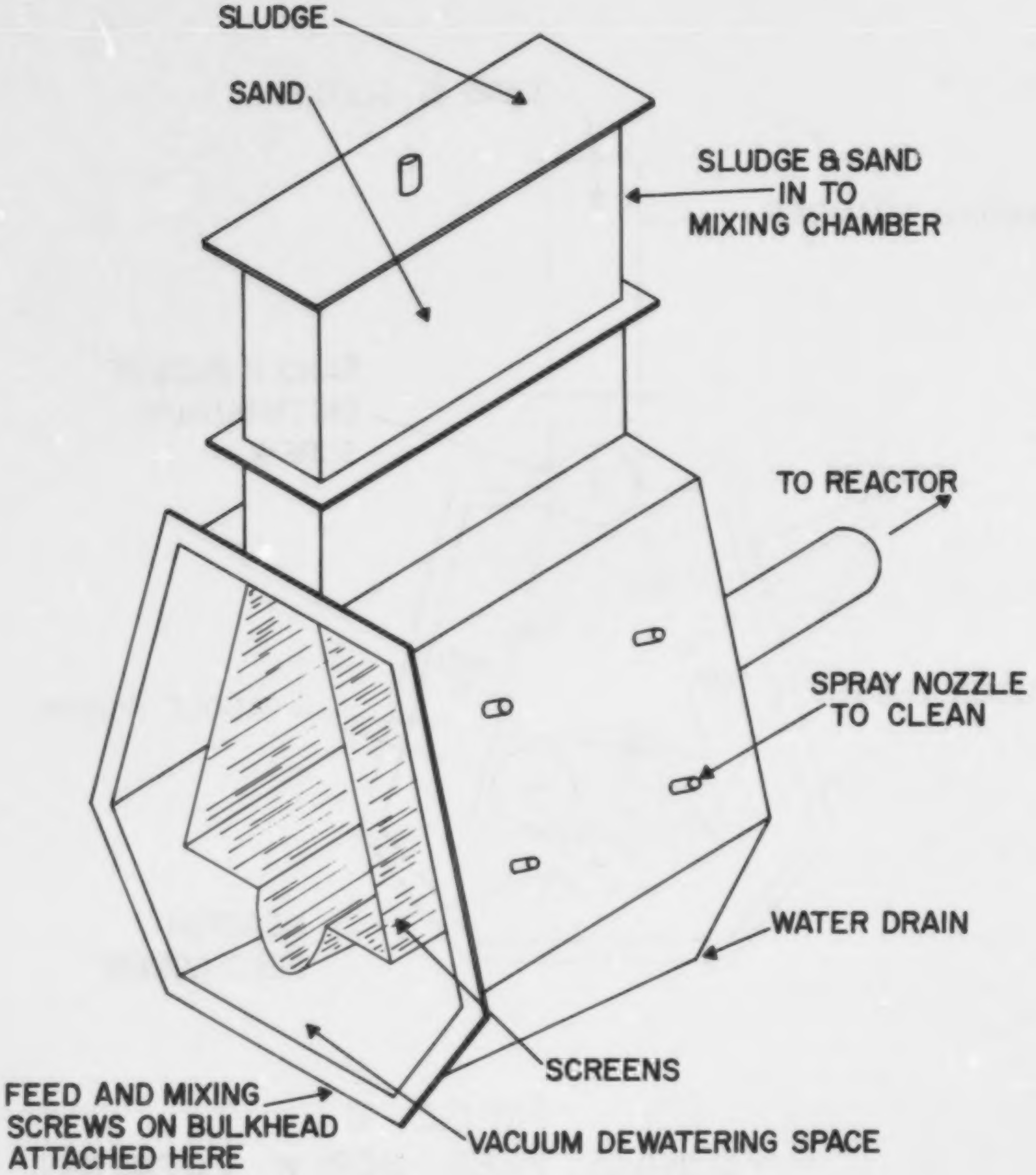

SLUDGE
SAND
SLUDGE & SAND IN TO MIXING CHAMBER
TO REACTOR
SPRAY NOZZLE TO CLEAN
WATER DRAIN
SCREENS
FEED AND MIXING SCREWS ON BULKHEAD ATTACHED HERE
VACUUM DEWATERING SPACE

FIGURE 7
SLUDGE DEWATERING
FILTER & FURNACE FEED SCREW CHAMBER
FREEHOLD TREATMENT PLANT

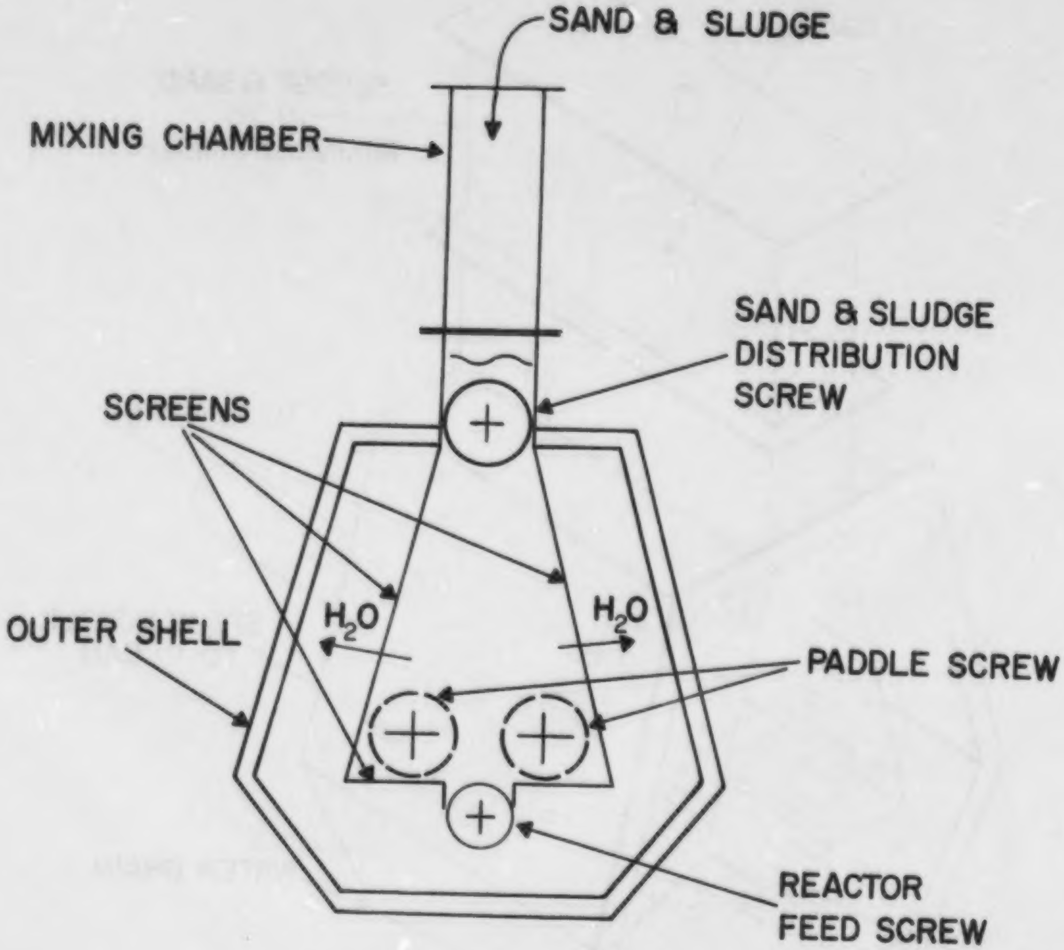

SAND & SLUDGE
MIXING CHAMBER
SAND & SLUDGE DISTRIBUTION SCREW
SCREENS
H_2O
H_2O
OUTER SHELL
PADDLE SCREW
REACTOR FEED SCREW

FIGURE 8
FURNACE FEED SCREW LOCATION
IN SLUDGE DEWATERING FILTER

FREEHOLD TREATMENT PLANT

screen in considerable quantity into the filtrate area. Dismantling of the filter showed that the gasket has been improperly placed.

c. Drive Chain Slippage

In early tests considerable difficulty was encountered with the drive chain jumping or slipping on some of the sprockets. This was attributed to the excessive loads caused largely by piling up of sand because of the improper arrangement of the mixing paddles. This condition was corrected by the adjustable paddle arrangement and also by installing larger sprockets on the lower mixing paddles, allowing operation at slower speeds with a greater wrap of chain around the sprockets.

d. Sludge Bypassing

In early tests considerable solids were found in the filtrate from the sand filter. This was believed to be caused by sludge bypassing sand at the top of the filter. In order to minimize this problem part of the filter screen was temporarily blanked off by tape in approximately the top six inches of filter area. Tests showed that it might be desirable to blank off the entire side area section of the screens. The unit was subsequently modified with an adjustable gasket arrangement so that any portion of the side screen could be blanked off.

These mechanical modifications consumed much of the time available for testing. Consequently, only very limited performance data was developed. Two primary observations evolved:

(1) Filtrate quality was very poor with suspended solids in the range of 0.3 to 1.5% when feeding sludge at concentrations of 1.2 to 3.0% solids.

(2) At high sludge feed concentrations, an almost impervious layer of sand and dewatered sludge mixture tended to built up against the side screens, preventing further filtration and also inhibiting downward movement of sand within the filter. This eventually caused a cavity to develop in the center of the filter, followed by discontinuation of discharge from the discharge screw, bridging above the screens, and inability to continue feed of sand and sludge to the top of the filter.

These difficulties were felt to be due to a combination of the mechanical problems outlined above and limited facilities at the test location. The unit was, therefore, installed at the Freehold Plant. It was believed that the modifications outlined above would permit satisfactory mechanical operation of the unit, and that more appropriate testing could be accomplished where the complete facilities and components of the filtration concept were available.

2. Incinerator

Early discussions on incinerator design revolved around the relative merits of insultated high temperature alloy construction versus refractory lined steel. It was decided that the most conservative design would be with refractory lining, and this approach was chosen. This choice dictated that the inside diameter of the unit be no less than two feet in order to allow access for a bricklayer into the unit.

The incinerator was designed with a superficial vapor velocity of approximately 2 ft./sec. For this velocity, the desired disengaging height is 8 ft. Because of space limitations the disengaging space provided was only 7 ft. above the expanded bed height of 4.9 ft. This causes a slightly greater loss of sand because of sand being carried over in the exit gas. The incinerator design was completed after the building and most of the equipment had been designed and laid out. It was therefore necessary to make much of the incinerator equipment fit the available space and to provide the capability of placing the reactor in the field, moving it around equipment already in place. This limitation dictated that the reactor be built and installed in three flanged sections. Installation of the incinerator required that the top section be suspended in place prior to setting and refractory lining the base and middle sections. This added to the reactor design, fabrication and installation time and cost.

Basic performance parameters for sludge incineration are shown in Table 3.

3. Fluidizing Blower

The blower to feed the incinerator was designed for 150 scfm with a discharge pressure of 10 psig.

4. Cyclone

The cyclone for separating solids from the incinerator offgas was specially designed for this service. It was fabricated of high temperature resistant alloy (RA 330). The design specifications were:

Incoming gas flow:
    Rate - Max. 250 scfm
    Temperature - 1600°F.
    Pressure - Approx. 6" Water
    Pressure Drop - 5" to 8" Water
    Collection Efficiency - 99% above 5 micron

TABLE 3

Incinerator Design Parameters

Parameters	Quantity
Bone Dry Sludge	13.65 #/hr.
Water	62.2 #/hr.
Ash	4.78 #/hr.
Air	77.6 scfm
Fuel Oil	2.36 gal/hr.
Operating Temp. (exit gas)	1600°F.
Diameter	2'
Static Bed Height	3.5'
Expanded Bed Height	4.9'
Sand Loss	ca. 1 #/hr.
Maximum Flue Gas Generation	110 scfm
Minimum Fluidization Velocity	0.255 ft./sec.

Dimensions:
Diameter - 22"
Height of Straight Side - 33"
Height of Conical Side - 55"
Nozzles: Gas Inlet 11" x 4.4"
Gas Outlet 11" dia.
Solids Outlet 6" dia.

5. Scrubber

The scrubber was designed to handle a maximum of 400 scfm of gases at 400°F. A Croll-Raynolds No. 8 x 8 Fume Scrubber with 24" Circulation Tank and Pump was selected.

6. Suction Blower

Incorporation of the Continuously Rengenerating Filter in the solids handling design required a suction blower to overcome the pressure drop from the incinerator through the cyclone and scrubber system in order to create a slightly negative pressure at the inlet to the filter. The blower was designed for 300-500 scfm at 3" mercury.

7. Carbon Rengeneration

The carbon regeneration concept in the overall design was based on interruption of the sludge incineration mode in the fluidized bed reactor for a period of about two days once each thirty days. During this two-day period, the accumulated spent carbon would be regenerated. The regeneration system was designed to interface with the treatment system on the following basis:

	mg/l	lb./day
BOD to Carbon Adsorber	40	16.7
BOD Residual	5	2.1
BOD Removed	35	14.6
Carbon Exhausted		29.2

29.2 lb/day x 30 days ÷ 48 hr. regeneration cycle = 18.25 lb/hr.

Preliminary carbon regeneration work was carried out using a 6" diameter laboratory fluidized bed reactor. General observations made as a result of this work are as follows:

a. Carbon recovery during steady state operations ranged from 85% to 100%.

b. Regenerated carbon quality as determined by relative capacity and iodine value was equivalent to virgin carbon.

c. The ratio of carbon to sand in the bed overflow increased with time.

d. For a given fluidization velocity, low initial static sand bed height reduced the percentage of sand in the overflow.

Initial design specifications for the carbon regeneration process were:

Carbon	18.25 #/hr.
Volatiles	9.125 #/hr.
Water	18.25 #/hr.
Fuel	3.05 gph
Air	64.6 scfm
Flue Gas	115.9 scfm
Temperature in Reactor	1400°F.

Later work showed that the water to carbon ratio in the spent carbon would be nearer 1.5:1 rather than the 1:1 originally assumed. This condition required more heat during the regeneration cycle than had been previously calculated; requiring several design changes.

a. The plenum temperature capability was increased from 1600°F. to 2000°F. to effect greater heat utilization.

b. The incinerator would be operated with limited air and with fuel injected directly into the bed.

c. A carbon feed screw was designed with a special drainage section to try to bring the water in the carbon feed to as low a value as possible.

8. Instrumentation

Major instrumentation in the fluidized bed reactor system included the following:

a. Plenum temperature recorder controller

b. Bed temperature recorder with high-low shut-off and alarm.

c. Upper bed temperature indicator.

d. Freeboard temperature indicator with high temperature shut-off and alarm.

e. Off gas temperature indicator with high temperature shut-off and alarm.

f. Scrubber inlet gas temperature indicator alarm.

g. Scrubber water temperature indicator alarm.

h. Scrubber water level control alarm.

C. AUXILIARY FACILITIES

Certain facilities in the plant were common to both the liquid handling and solids disposal processes.

1. Oil Tank

2,000 gallon capacity, steel: Fuel was needed for three purposes: incinerator, emergency generator, and space heaters. It was originally thought that natural gas could be used for the space heater and the emergency generator, but it developed that no natural gas would be available in the community. Since fuel oil was to be used for the fluid bed incinerator, the generator and heater were changed to fuel oil. A 2,000-gallon tank size was selected. This was located outside, underground with a level indicator showing inside the building.

2. Oil Pump

A gear pump with a capacity of 0.23 gpm against a head of 276 ft. of fluid was selected.

3. Air Compressor and Receiver

10.8 cfm, 100 psig, 3 HP; with 240 gallon receiver. The air compressor was sized to handle instrument and other requirements, with the largest instantaneous use being flushing of the magnetic filter. The oversize receiver was specified to handle this large but infrequent load.

4. Emergency Generator

A 50 kilowatt emergency generator was selected. This would supply power to the liquid handling equipment in the plant in case of any loss of power. Because of the large size of the blower motors on the incinerator, and because sludge storage capacity was provided, it was decided that it was not economical to supply emergency power for the incinerator section. In order to avoid overloading the emergency generator by the high starting current on motors, a timer was installed to sequence the startup of some motors.

5. Instrument Air Drier

A refrigeration type air drier was provided to insure a source of dry air for instrumentation.

Inlet Flow:	25 scfm
Inlet Pressure:	100 psig
Outlet Moisture:	-10°F. dewpoint at atmospheric prssure

6. Sump Pump

One-half HP open impeller, with pressure switch actuation; 50 gpm at 20 ft. head.

SECTION V

PHASE 2. CONSTRUCTION

A. INTRODUCTION

One objective of this project was to locate the treatment plant within the housing development subdivision which it would serve. Since a portion of the development plot bordered on the Manasquan River the plant site was selected for convenient discharge into this receiving stream. Figure 9 is a vicinity map showing the location of the treatment plant within the subdivision. A more detailed view of the plant site is depicted in Figure 10, showing the adjacent playground area, location of the lift station which delivers raw sewage to the plant, and the point of effluent discharge into the Manasquan River.

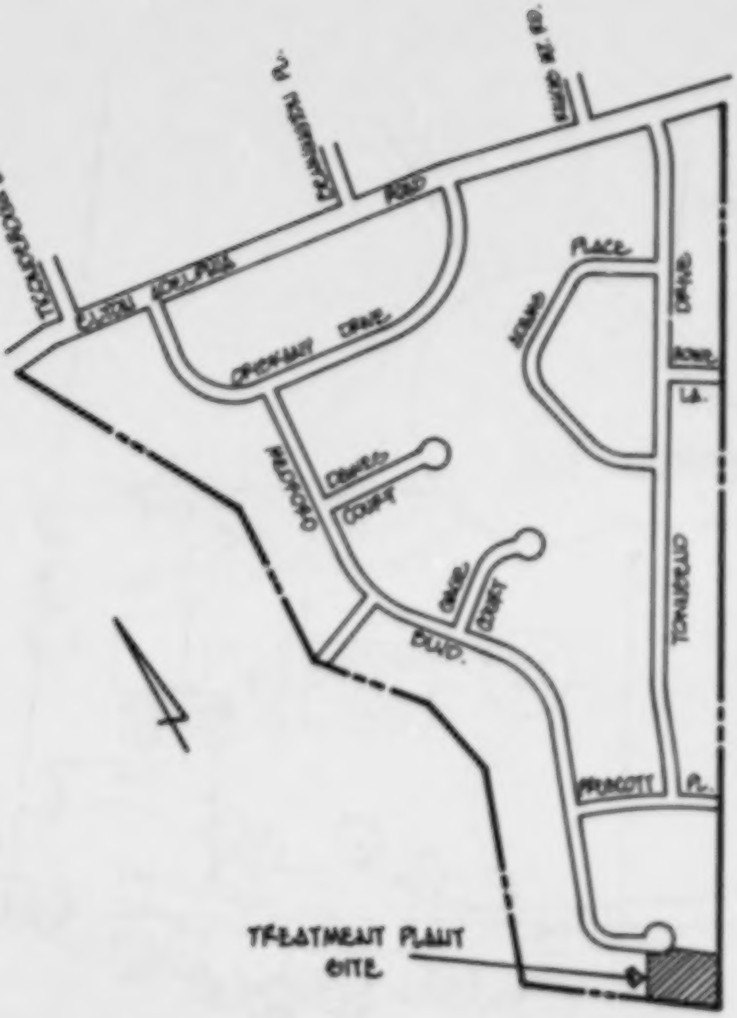
Construction of the treatment plant was carried out under a number of contracts and subcontracts for equipment and construction services. Coordination of this phase with other aspects of the total project was essential in order to meet the schedule objectives. As a result some overlap of work was necessary, leading to a few problems and inefficiencies which would not have been encountered otherwise.

For example, the treatment plant was scheduled to be the only route for disposing of sewage from the housing development. It was necessary, therefore, that the plant be operational, at least in the liquid handling process, before the first house in the development was completed and occupied. This requirement in turn dictated early completion of construction design before part of the process engineering design was completed.

A critical path schedule was developed for Phase 1 and 2 of the project as a guide to the most efficient route to completion. In general, the construction phase was carried out with no major insurmountable problems although some annoyances and delays did develop.

B. CONSTRUCTION DETAILS, PROBLEMS AND SOLUTIONS

The balance of this section deals principally with difficulties which were encountered during construction, solutions to these problems, and recommendations, where appropriate, for avoiding similar pitfalls in other projects of this type.



ROAD
PLACE
DRIVE
LA.
COURT
BLVD.
COURT
PRESCOTT
PL.
TREATMENT PLANT
SITE

FIGURE 9
VICINITY MAP
FREEHOLD TREATMENT PLANT
TOWNSHIP OF FREEHOLD, MONMOUTH CO., N.J.

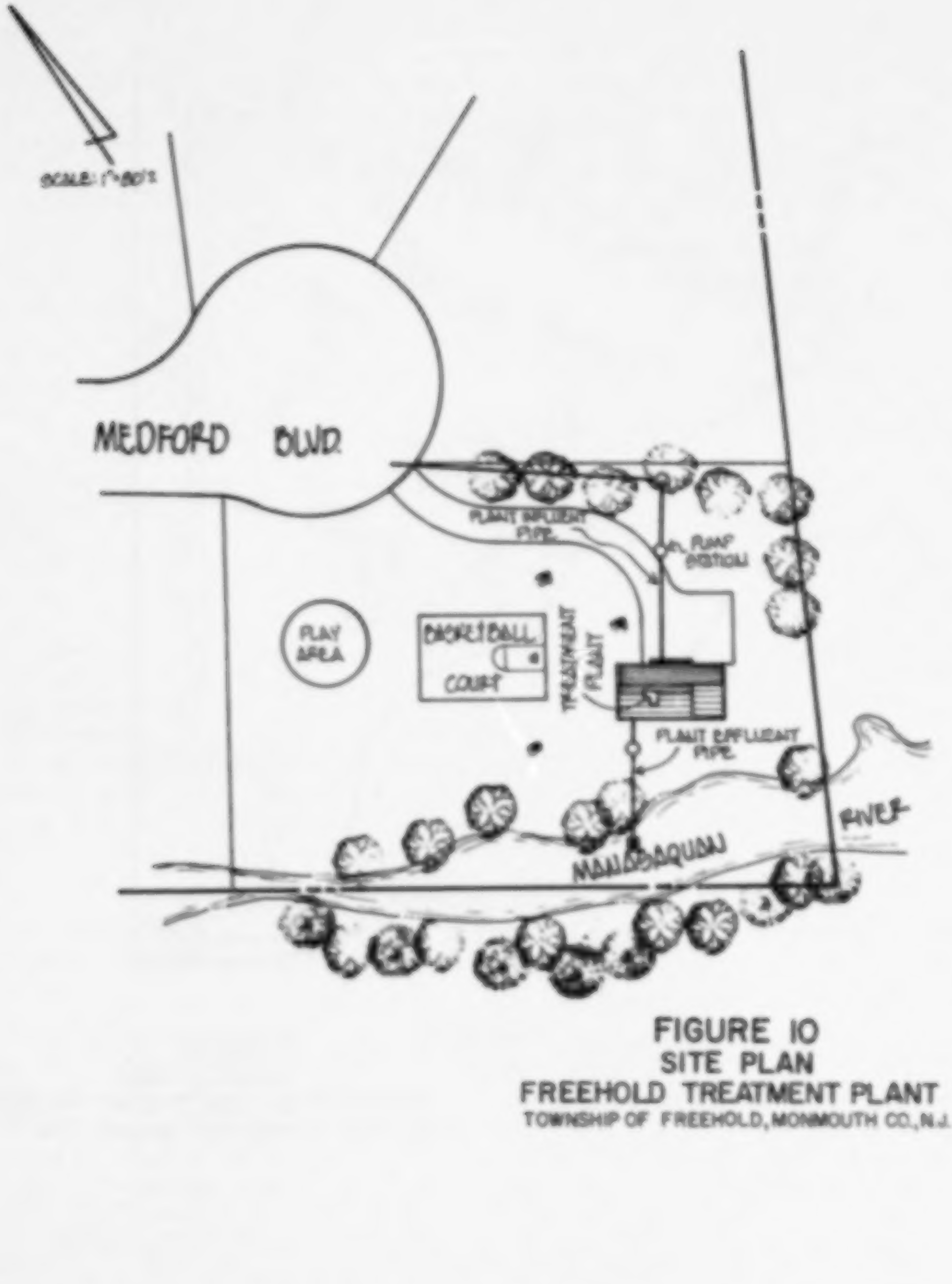

SCALE: 1"=80'±
MEDFORD BLVD.
PLANT INFLUENT PIPE
PUMP STATION
PLAY AREA
BASKETBALL COURT
TREATMENT PLANT
PLANT EFFLUENT PIPE
MANASQUAN
RIVER

FIGURE 10
SITE PLAN
FREEHOLD TREATMENT PLANT
TOWNSHIP OF FREEHOLD, MONMOUTH CO., N.J.

1. Building Selection

It was originally planned to enclose the treatment plant in one of the four housing models offered in the development. The first choice was a so-called high ranch style, a two-story house, *chosen primarily because it offered the largest envelop. It was* later decided, however, to change to a single-story Cape Cod style, the Phoenix, for a lower profile and to blend into the neighborhood. This house had external dimensions of 24'8" x 50'. In order to utilize the Phoenix house a basement 13 feet deep had to be incorporated in the design. Since economics dictated that no rock excavation would be done, the floor of the basement was built on bed rock. This put the ground floor of the house several feet above the existing grade, thereby requiring extensive grading around the house.

Although the building selection proved generally satisfactory some improvements in cost and equipment arrangement would likely have resulted from a slightly larger (10-20%) envelop and construction on a single floor layout.

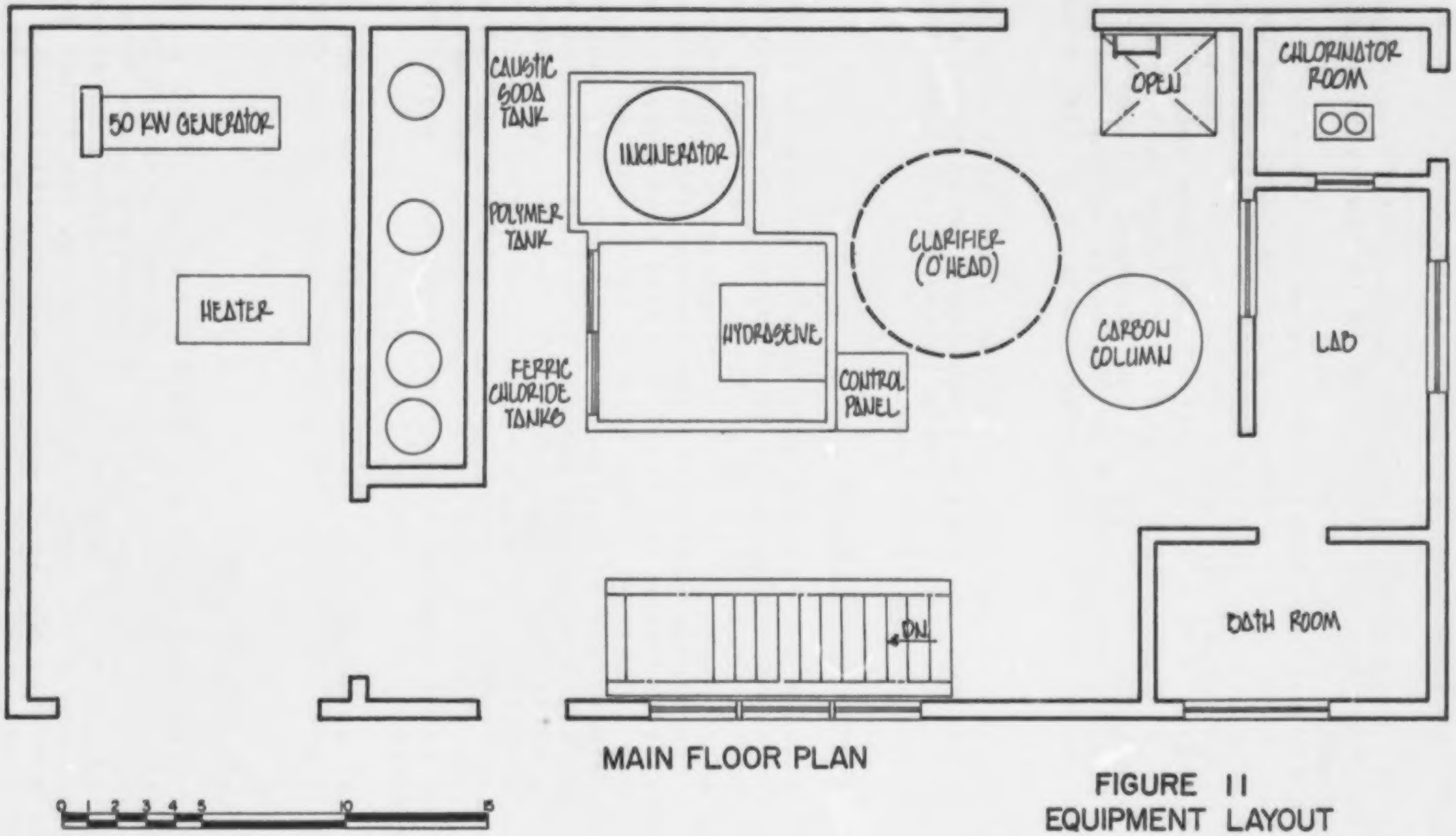
2. Arrangement and Layout

As noted above, it was necessary to proceed with the building design and construction while development work and equipment engineering design were still in progress. As a result, the size and optimum arrangement of several major pieces of equipment, primarily in the sludge dewatering and incineration sections, had not been established when the final building configuration was selected. This situation, compounded by space limitations, made the physical dimensions of equipment a very important consideration. In several cases extensive engineering effort was expended to satisfy the simultaneous requirements of field installation, accessability for operation and future maintenance, relationship to other equipment, and available space. The principal problems involved carbon transfer piping layout, clarifier location and design, carbon configuration and arrangement, and provision for a small office-laboratory space.

Final arrangement of major equipment in the plant is shown in Figures 11 and 12. Figure 13 is a photograph of the completed building.

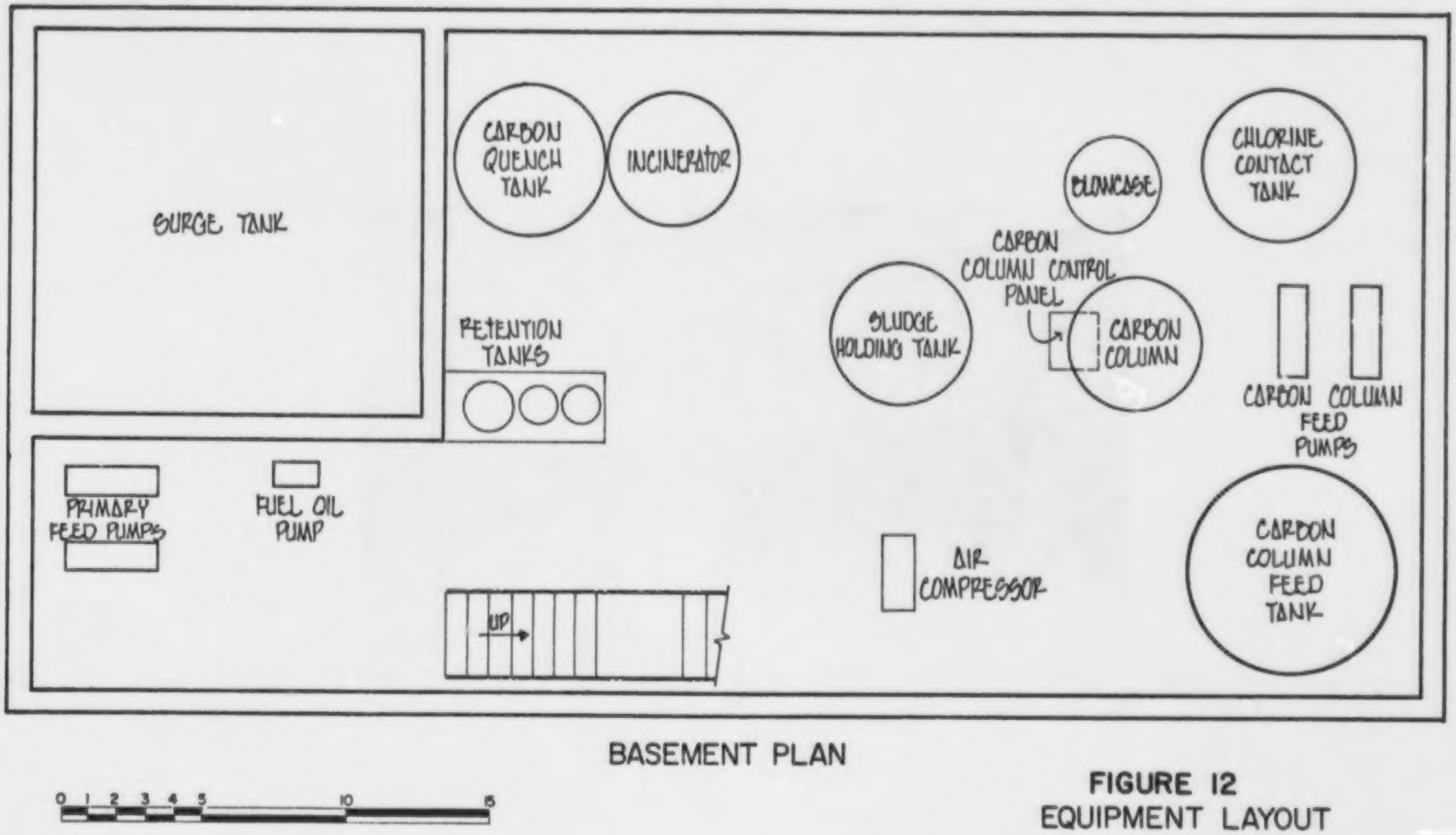
3. Access and Services

The treatment plant was the first structure to be built in the development, and initial construction work preceded the building of the permanent road to the plant. A temporary acess from a nearby road was bulldozed to the plant site. This was not an all-weather road, however, and there were many days during the Winter and Spring when access was very difficult or impossible, particularly for large trucks and cranes, needed for placing heavy equipment. An unusually mild winter contributed to the difficulty, as the ground did not


50 KW GENERATOR
HEATER
CAUSTIC SODA TANK
POLYMER TANK
FERRIC CHLORIDE TANKS
INCINERATOR
HYDRASEIVE
CONTROL PANEL
CLARIFIER (O'HEAD)
CARBON COLUMN
OPEN
CHLORINATOR ROOM
LAB
BATH ROOM
DN
0
1
2
3
4
5
10
15

MAIN FLOOR PLAN

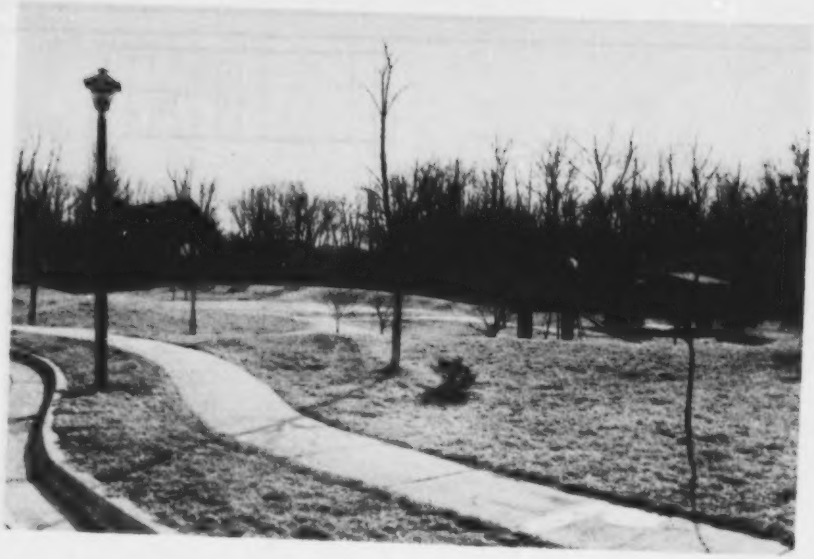
FIGURE 11
EQUIPMENT LAYOUT
FREEHOLD TREATMENT PLANT
TOWNSHIP OF FREEHOLD, MONMOUTH Co., N.J.


SURGE TANK
CARBON QUENCH TANK
INCINERATOR
RETENTION TANKS
PRIMARY FEED PUMPS
FUEL OIL PUMP
UP
SLUDGE HOLDING TANK
CARBON COLUMN CONTROL PANEL
BLOWCASE
CARBON COLUMN
CHLORINE CONTACT TANK
CARBON COLUMN FEED PUMPS
CARBON COLUMN FEED TANK
AIR COMPRESSOR
0 1 2 3 4 5 10 15

BASEMENT PLAN

FIGURE 12
EQUIPMENT LAYOUT
FREEHOLD TREATMENT PLANT
TOWNSHIP OF FREEHOLD, MONMOUTH Co., N.J.

FIGURE 13
PLANT EXTERIOR



FREEHOLD TREATMENT PLANT

freeze in the area. This problem was finally solved by hauling in sufficient rock to stabilize the access road.

No permanent electric service was provided during the construction phase, and a portable generator was required to furnish power for hand tools and other equipment.

The driveway to the treatment plant, completed in the latter stage of the development, was of the same type as in the adjacent home construction. It was not heavy enough for the large trucks delivering equipment and materials to the plant. In addition, the driveway was designed with an "S" curve which made it very difficult for a truck to back into the area. The net result was considerable damage to the lawn area and driveway. The driveway was ultimately straightened and strengthened.

4. Miscellaneous

The surge tank was constructed of reinforced concrete. Leaks were encountered when the tank was initially filled with water for testing. Two patching attempts, requiring draining and refilling the tank, were needed to completely eliminate the leaks.

The building contractor misplaced the Hydrasieve discharge opening in the poured concrete floor. It was then necessary to cut out and repour part of the floor.

During the electrical check out phase an electrician inadvertently connected 100 volt AC power to the low voltage DC circuits in the instrument panel. This accident damaged a number of alarm systems which had to be replaced.

Several delays caused a number of problems in other phases of the construction. The most serious of these were:

a. The access problem referred to above.
b. Overlap of development and design of the solids disposal system with plant design and construction as discussed earlier.
c. Slow completion of the building, requiring double handling of delivered equipment.
d. Late delivery of the carbon adsorption system, interfering with the scheduling of piping and electrical work.

SECTION VI

PHASE 3. OPERATION AND EVALUATION

A. LIQUID HANDLING PROCESS

1. Plant Startup and Break-In: October 1972-June 1973

a. General Description

The Woodgate Farms wastewater treatment plant uses physical-chemical technology, patterned after the pilot system described in Section IV.

Large solids are removed by screening; fine suspended solids and some dissolved solids are removed by chemical flocculation, settling, and filtration; and dissolved organic matter is removed by adsorption on granular activated carbon. The effluent is disinfected and discharged to the river. The settled sludges are converted to an inert ash by incineration in a fluidized bed reactor. The plant is automated for operation with minimum attention. A schematic Process Flow Diagram is shown in Figure 14.

b. Process Description

(1) Lift Station

Gravity mains in the community deliver raw domestic sewage to the lift station which is near the treatment plant. This station is a vacuum lift type unit with two centrifugal sewage pumps, check valves, a vacuum chamber and pump, level control floats, and electric starters and controls. The vacuum pump and chamber keep the centrifugal pumps primed by not allowing sewage to drain back to the well when a pump shuts down. The electrical switching gear alternates pump operation, thus equalizing wear. The pumps can be operated automatically or manually. The station is equipped with an automatic temperature-controlled heater to prevent freezing during cold weather.

When the sewage level in the wet well rises to a predetermined point set by float No. 1, one of the lift pumps will start and run until the level is pumped down to the low set point of the float, then it will shut off. If for some reason, such as a very high inflow or a defective pump, the level continues to increase, float control No. 2 will start the second pump. If the

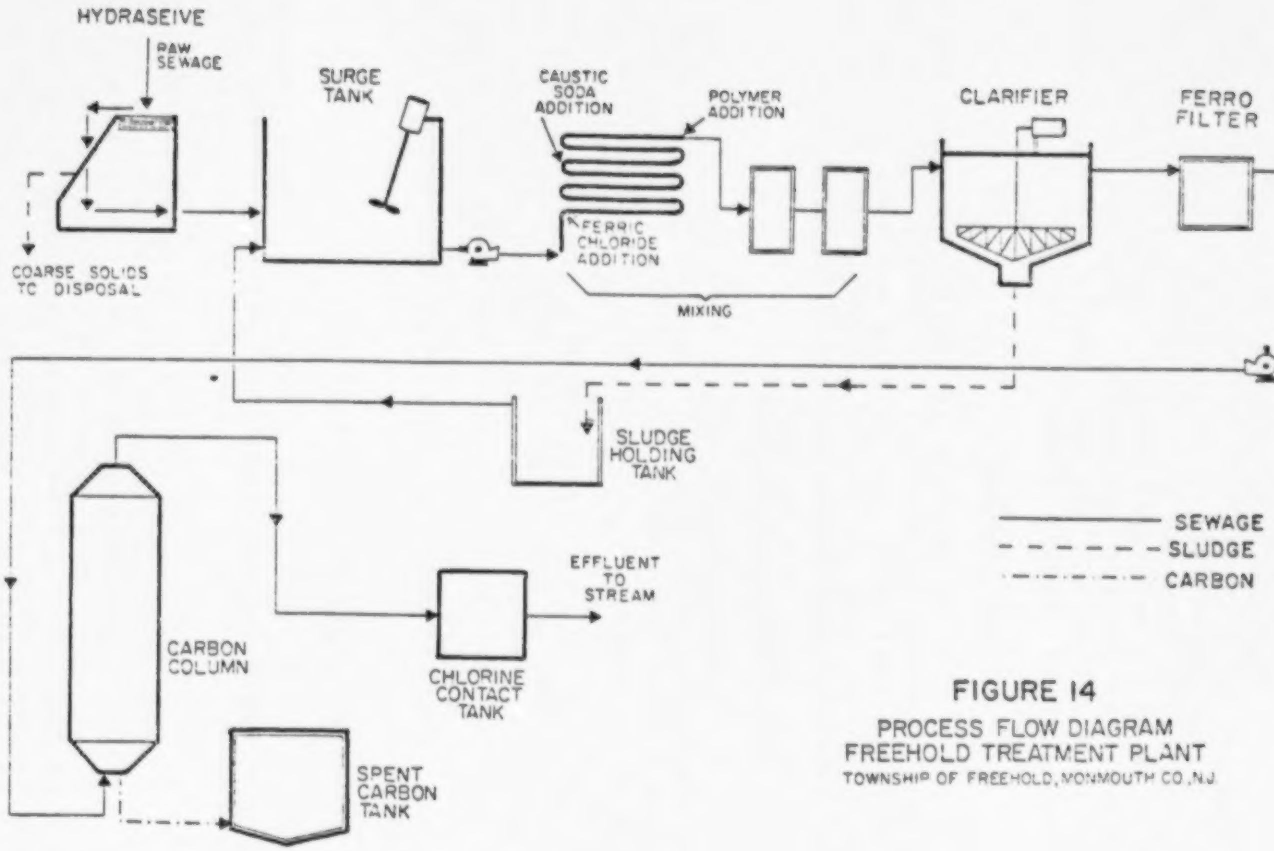

HYDRASEIVE
RAW SEWAGE
COARSE SOLIDS TO DISPOSAL
SURGE TANK
CAUSTIC SODA ADDITION
FERRIC CHLORIDE ADDITION
POLYMER ADDITION
MIXING
CLARIFIER
FERRO FILTER
CARBON COLUMN FEED TANK
SLUDGE HOLDING TANK
CARBON COLUMN
CHLORINE CONTACT TANK
EFFLUENT TO STREAM
SPENT CARBON TANK
SEWAGE
SLUDGE
CARBON

FIGURE 14

PROCESS FLOW DIAGRAM
FREEHOLD TREATMENT PLANT
TOWNSHIP OF FREEHOLD, MONMOUTH CO., N.J.

level increases above this point, float No. 3 will activate a "High Wet Well Level" light on the process control panel and an alarm will sound.

(2) Primary Screening

Raw sewage from the lift station is pumped to the distribution box of a Bauer "Hydrasieve", a 3 ft. wide sloping wedge wire screen with 0.020 inch openings between the wedge wires. As the sewage flows down the screen, solids greater than 0.020 inches in size are separated from the liquid and slide into a collection hopper for further processing. The liquid flows through the screen to a collection pan and then by gravity to the Surge Tank. The screen can handle flows of 150 GPM.

(3) Surge Storage

The Surge Tank serves as a storage and equalizing vessel between the highly variable raw sewage influent flow and the subsequent operations which are carried out at a constant rate. The capacity of the tank is 15,680 gallons. It is constructed of reinforced concrete and is equipped with an agitator.

(4) Process Flow Control

The screened liquid from the Surge Tank is pumped through a chemical mixing-retention loop to the clarifier by one of two Process Feed Pumps rated at 50 GPM each. The flow rate is controlled by an automatic valve in the process line actuated by the Process Flow Recorder-Controller on the control panel. This instrument will automatically maintain a present flow when the switch on the front of the controller case is in "Auto" position, and the switch on the side is in the "Local" position. It will increase the flow as the surge level increases and decreases the flow as the surge level descreases when the switch on the side is moved to the "Remote" position; however, this method of control was not used.

(5) Chemical Treatment

As the screened liquid is pumped through the mixing loop enroute to the Clarifier, three chemicals are added continuously.

(a) Ferric Chloride is added to react with the phosphorus to form insoluble ferric phosphate. It also forms ferric hydroxide, a gel-like material which acts as a coagulant. The ferric chloride is continuously injected into the process line downstream of the process flow control valve at a tee by an electronically operated chemical metering pump. This pump can be controlled automatically by a turbidity sensor, or it can be controlled manually. Ferric chloride is received in 50 gallon drums as a 40% solution by weight of $FeCl_3$ in water. When it is transferred to the feed tank, one gallon of 40% ferric chloride is diluted with one gallon of water for greater accuracy of metering into the system.

(b) Sodium Hydroxide, added for pH control, is continuously injected into the process stream halfway through the retention loop, at a 180° bend, by an electronically controlled chemical metering pump similar to the ferric chloride pump. The pump is regulated by a pH Recorder-Controller which is set to control the pH of the process stream at the desired value; usually about 7.0. Sodium Hydroxide (caustic soda) is received in 50 gallon drums as a 50% solution by weight. When it is transferred to the feed tank, one gallon of 50% caustic is diluted with one gallon of water.

(c) Flocculant is added downstream from the caustic addition point at the end of the retention loop. A 0.2% solution of Hercofloc 836.2, a polyelectrolyte, is used. It induces efficient flocculation, incorporating the suspended sewage solids and ferric precipitates into large, rapidly settling flocs. The flocculant is injected by a chemical metering pump, manually controlled but otherwise similar to the ferric pump. Hercofloc 836.2 is received in 50 lb. bags as a white powder. The 0.2% solution is prepared by dissolving the calculated amount of powder in hot water using an eductor and agitator.

Kenics mixers were originally installed in the pipelines downstream of each chemical feed point but were later removed because of excessive plugging.

(6) Clarification

The flocculated sewage enters the clarifier via a central downflow feed well, then flows up through a sludge blanket. The clarified liquid leaves the clarifier peripherally through "V notch slots to a collection trough. A continuous slip stream of the clarifier effluent runs through a turbidity meter. The results are continuously recorded as JTU by the Turbidity Recorder on the control panel. If the turbidity rises above a preset level (8.4 JTU) the clarifier effluent is automatically recycled to the Surge Tank and the High Turbidity alarm sounds.

Sludge discharge is assisted by a slow moving rake (0.3 RPM). Ultrasonic type fixed probe sludge level detectors monitor the sludge level and automatically start or stop the Clarifier Sludge Pump, as required to maintain the predetermined sludge level desired in the clarifier. The sludge is pumped to the Sludge Holding Tank.

(7) Filtration

The Clarifier effluent flows by gravity through a magnetic filter to the Carbon Adsorber Feed Tank. Magnetic iron oxide slurry is injected into the effluent stream enroute to the filter. The magnetic iron oxide is received in 50 lb. bags. It is slurried in water at a concentration of 0.5% by weight and pumped by a chemical feed pump similar to the flocculant pump.

The magnetic filter (Ferrofilter) consists essentially of a water cooled shell with copper coils which, when electrically energized, produce a strong magnetic field. The center of the shell through which the liquid flows is packed with iron grids. The iron oxide added to the clarifier effluent stream attaches to or associates with any floc particles leaving the clarifier, then is trapped and holds the floc in the magnetic field of the filter. The magnetic field is turned off, and the filter washed via an automatic valve arrangement to the Surge Tank at regular intervals as determined by timers in the control panel. These timers can be set to obtain the desired frequency as well as duration of air scour and wash.

(8) Carbon Adsorption

The Ferrofilter effluent flows by gravity to the Carbon Adsorber Feed Tank which has a capacity of 2,000 gallons. The tank is normally nearly full, the level controlled by a float operated valve, because water from this tank is used to cool the Ferrofilter jacket and as the supply for the incinerator off gas spray nozzles and scrubber. Should the level in the Feed Tank get too high it will overflow to the basement sump and be pumped to the Wet Well. Should the level drop too low, a pressure switch opens an automatic valve to add city water. An approved backflow preventer valve was in the water line.

The Adsorber Feed Pumps, rated at 50 GPM each, pump effluent from the Feed Tank through the Carbon Adsorber to the Chlorine Contact Tank. The float operated valve, described under Item 8, in the discharge line from the Adsorber will open or close as required to maintain the proper level in the Feed Tank.

The Carbon Adsorber utilizes granular activated carbon to adsorb the dissolved organic material in the effluent stream. The water enters the bottom through a screen, moves up through the carbon and leaves at the top via four Johnson well point wedge wire screens. Part of the water returns to the Feed Tank as required to maintain the proper level. The remainder passes through a flow meter and enters the Chlorine Contact Tank. Periodically, a small amount of carbon is removed from the base of the column. The carbon in the column is washed with water and an air scour, and then the carbon is put back into the top of the column.

(9) Disinfection

The effluent stream is disinfected with chlorine in the Chlorine Contact Tank. The chlorine is supplied from a cylinder in the Chlorine Room and is added via an eductor driven by a circulating pump. The chlorinated effluent is discharged to the Manasquan River.

(10) Sludge Storage

As described in the Clarification section, sludge is pumped from the Clarifier and stored in the Sludge Holding Tank. The sludge pump is controlled by the sludge level detectors in the Clarifier which are set to turn the pump on at a predetermined upper level and turn it off at a lower level so as to maintain a sludge blanket. This is designed to produce a sludge ranging from 5 to 8% total solids. The speed of the pump can be manually changed to vary the pumping rate as may be required.

The sludge is pumped to the Sludge Holding Tank where it is stored. Hydrasieve solids are also pumped to this tank from the Hydrasieve Hopper via the grinder pump. This is a manual operation performed daily. The sludge in the Holding Tank thickens by settling. Periodically, clear liquid can be decanted from the tank via the decant nozzles. The original design provided for sludge to go through the sand filter into the Incinerator.

Sludge is pumped to the Incinerator by the Incinerator Sludge Feed Pump. This pump is controlled by an indicator-controller on the Incinerator panel which shuts the pump off if the bed temperature drops below the set point and restarts the pump when the temperature rises above the set point provided the pump switch is on "Auto".

(11) Incineration

The sludge is destroyed by fluid bed incineration at 1400-1500°F. which produces a sterile ash and odorless flu gases. The details of the operation of the fluid bed reactor system are contained in the Solids Disposal section.

c. Problems, Solutions, Modifications

(1) Equipment Checkout

During August and September, 1972, as plant construction was nearly complete and prior to start up, the pipe lines were flushed out and checked for leaks, all electric motors tested for proper rotation, and the tanks and vessels washed out and calibrated. The reinforced concrete Surge Tank leaks were repaired.

The four chemical feed pumps were tested and calibrated with water. The control instruments were tested and adjusted. The Clarifier effluent turbidity meter was calibrated and the recorder-controller set point set to divert the effluent flow back to the Surge Tank when the turbidity exceeded 8.4 JTU.

The Clarifier weir was leveled while the unit was full of water so as to obtain a uniform overflow distribution. The Carbon Adsorber was loaded with granular activated carbon by preparing a carbon-water slurry in the Quench Tank and educting it to the Adsorber.

The laboratory was cleaned. Major laboratory equipment was installed and calibrated as needed. Analytical procedures were checked using blanks and known standards.

(2) Startup Activities

On September 27, 1972, sewage from an adjacent community was pumped to the Lift Station via a force main connection. The Lift Station automatically pumped this sewage to the Surge Tank and treatment was begun. During the first three months sewage from this force main was admitted as needed to maintain plant operations for tests and adjustments. The use of force main sewage was discontinued in December 1972. At a later date the valve in the connecting line was removed and the line was plugged.

From the initial startup until late December 1972, the plant was operated intermittently as required to treat the effluent sewage. Starting in late December, the flow increased so that continuous, around-the-clock operations could be sustained. Initially, the plant was checked several times during the night. As start-up problems described in the following sections were solved, the frequency of operator attendance was reduced. By the latter part of the start-up period, the plant was running unattended about sixteen hours a day during the week and about twenty-one hours a day during weekends and holidays.

(3) Problems

The problems encountered during the start-up period, together with changes in equipment or procedures to solve the problems, are described according to the unit operations involved and not chronologically.

(a) Lift Station

The lift station failed at irregular intervals during the start-up period. The cause of the failures and the solutions were as follows:

[1] The level control float would swing over and hang up on carbon and other solids accumulated at one side of the wet well, allowing the liquid to be pumped down to the level of the priming tank suction pipe. Air would enter the pipe and the priming tank thus would lose the vacuum lift which keeps the sewage pumps primed. The surge of air into the vacuum priming tank would jam the vacuum pump surge float closed so the vacuum could not evacuate air from the priming tank to return the vacuum lift to normal operation.

The carbon and other solids were removed from the wet well and the floats and float guides readjusted. In addition, screen baskets were installed on the carbon column and Quench Tank drains to catch any carbon so it can be reused. A longer surge chamber supplied by the manufacturer was installed on the vacuum pump.

[2] After about six months of operation, the high humidity of the lift station caused corrosion of the switching gear contacts so that arcing occurred and the pumps would sometimes fail to start. The contacts were cleaned and a regular maintenance and inspection procedure was initiated.

[3] Small sticks, rags, and similar objects hung in the check valves, preventing them from closing when the pumps stopped. Thus, air flows back through the line and the pump prime was lost. Changing the check valve spring tension did not solve the problem. Submersible pumps were installed at a later date which solved the problem.

(b) Hydrasieve

The initial openings between the wedge wires were 0.020 inches. The screen had to be cleaned several times a day because of blinding with grease. The screen was exchanged for one with 0.040 inch openings. Performance of this screen was better, but it still blinded during the unattended periods, especially during late evening hours. A hot water spray wash was installed complete with timers to control the duration and frequency of the wash. Experience has proven that satisfactory operation is obtained with a spray frequency of 1 1/2 to 2 hours and a spray duration of 2 minutes. In addition, the Hydrasieve is manually cleaned with a brush and cleanser each morning. This hot spray wash and cleaning procedure has resulted in good Hydrasieve performance.

(c) Surge Tank

The Surge Tank is vented to the Hydrasieve enclosure via the influent line from the Hydrasieve and to the Chlorine Contact Tank via the emergency overflow line from the Surge Tank. There is also an access manhole from the Storage Room above the tank. The tank is agitated, but not aerated. Odor has never been a problem.

Periodically, during the nine-month starting period, problems were encountered with floc floating in the Clarifier. After considerable study, it was noted that this problem occurred when the Surge Tank level was below the 21% mark. It was determined that at levels below about 21% the agitation was such that air became entrained and dissolved in the liquid, which was then pumped through the chemical treatment loop to the Clarifier. The air, disengaging in the Clarifier, caused some flocculated solids to float instead of settling.

The Surge Tank Level sensing system was modified to control the agitator. When the Surge Tank level drops to 25%, the controller stops the agitator. When the level increases to 30%, the controller restarts the agitator. This procedure solved the problems of air entrained floating solids in the Clarifier. No problem was encountered with solids that settle in the Surge Tank during periods when the agitator is off.

(d) Chemical Treatment

The ferric chloride, sodium hydroxide, flocculant, and magnetic iron oxide were each added to the process flow stream by Precision Control Products electronic pulser, diaphragm type, chemical metering pumps. Just downstream from each injection point, two-inch diameter Kenics static mixers were used to rapidly and completely mix each chemical with the process stream. To avoid a pulsed flow of chemical, small diameter tubing and an accumulator were installed between each pump and the process line. Early in the start-up period, fifty foot lengths of capiliary tubing were inserted between the accumulators and the injection points. This resulted in a fairly uniform flow of chemical.

The static mixers seemed to be very effective in mixing the chemicals with the process stream; however, plugging of the elements with fiberous material from the process sewage stream was a serious problem. Once a week the plant had to be shut down to remove and clean the three mixers (ferric chloride, sodium hydroxide, and flocculant) upstream from the Clarifier. This required three to four hours of operator time. A strainer was installed upstream from the mixers. The mixer cleaning frequency was reduced from once a week to once every two weeks; however, the strainer had to be cleaned daily, a simple task requiring about ten minutes of operator time. The two-week cleaning frequency was unacceptable; therefore, the static mixers were removed and the chemicals introduced at tees in the process line. Clarification and effluent quality were satisfactory, but chemical consumption increased.

The original ferric chloride, flocculation and magnetic iron oxide feed tank agitators were stainless steel powered by 1/20 HP motors. They were used to dilute and mix the chemicals. These agitators were too small and lacked sufficient power to adequately mix the chemicals in a reasonable time. The ferric chloride agitators were removed. An air sparge was used to mix the 40% as received solution with water prior to feeding to the process stream. A portable 1/4 HP mixer was used to prepare the flocculant solution. An air sparge was installed and ran continuously to improve the agitation of the magnetic iron oxide slurry to keep it in suspension.

(e) Clarifier

The original design of the Clarifier provided for the process flow to enter the center well in a downward direction parallel to the rake shaft, and to discharge laterally against the inside wall of the center well through a tee installed at the end of the inlet pipe about one foot below the surface. During the initial start-up period, it was found that at flows of about 20 GPM and higher the clarifier was plagued with erratic operation and periods of high clarifier effluent turbitidy. This problem appeared to be aggrevated by operation of the clarifier rake, which had a speed of 0.67 rpm. It was soon discovered that the tee had not been installed on the inlet line, and that the influent flow was producing a jet-like disturbance of the sludge blanket. Installation of the tee and reduction of the rake speed to 0.3 rpm eliminated this type of clarifier sludge problem.

In the fabrication of the clarifier a conical wooden block was installed at the bottom just below the rake hub. Its purpose was to minimize the accumulation of sludge in what might be a dead spot. This cone began to disintegrate after about six months of operation, and pieces of wood occasionally plugged the sludge discharge nozzle. The remains of the cone were finally removed. Sludge accumulation, if it occurs, on the flat center section under the rake hub has not been determined to be a problem.

The Clarifier effluent leaves via a circular V-notch weir and collection trough to a float chamber and from there by gravity through the Magnetic Filter to the Adsorber Feed Tank. A slipstream of about two liters per minute flows by gravity from the float chamber to a continuous turbidity meter.

The turbidity controller is normally set to divert the flow back to the Surge Tank if the Clarifier effluent is higher than 8.4 JTU. The float was designed to regulate a valve down-stream of the magnetic filter so as to maintain a constant level in the float chamber, thus assuring a representative slipsteam flow to the turbidimeter. Despite frequent adjustments to the float control system, the water level fluctuated to the extent that periodically air bubbles would be drawn into the turbidimeter slipstream giving erroneous readings and control. The float control system was removed and a short standpipe inserted in the discharge line. This eliminated the air bubbles, yet provided a slipstream typical of the Clarifier effluent.

Erratic sludge level control was a reoccurring problem during the first six months of operation. Frequently, the ultrasonic sludge level controller would fail to start or stop the sludge pump as required to control the sludge level in the Clarifier. Thus, the sludge level would get too high, resulting in solids carryover; or too low, resulting in loss of the sludge blanket leading to less efficient clarification and low sludge solids in the sludge hold tank. Representatives of the manufacturer, working with plant operating personnel, determined that the erratic operation was caused by misalignment of the Sensall probes, which had been supplied on an adjustable mount during the development phase. Factory aligned probes in a rigid assembly were then installed and the controls adjusted. Sludge level control has been good since this change was made.

(f) Clarifier Sludge Pump

The Clarifier sludge pump was originally located on the basement floor under the Sludge Tank. The pump was a positive displacement type with an eccentric rotor and a rubber-lined stator. This pump soon failed because of wear between the sides of the rotor and the stator. It was replaced with a stainless pump with a flexible gear-like rotor made of rubber. The rubber rotor failed after a few weeks of service. A Moyno, type FS, progressing cavity pump with a chrome plated stainless rotor and synthetic rubber stator was installed. The high head, about 23 feet, on the suction caused frequent collapse of the synthetic rubber stator. The pump was moved to the Hydrasieve room reducing the suction head to about 7 feet. The operation and maintenance history of this pump has been good. The stator needs to be replaced at three-month intervals. The rotor lasts about nine months.

Considerable difficulty was encountered in handling sludge containing five to eight weight percent solids in small diameter lines, i.e. lines less than 1.0" diameter. Steel pipe was replaced with reinforced synthetic rubber hose, using the minimum number of valves and fittings to avoid points where plugging might occur. In addition, a procedure was established to flush the sludge line with water each day. These changes reduced plugging problems to infrequent intervals.

(g) Carbon Adsorption System

In the initial loading operation of the carbon adsorption column, some difficulty was experienced in moving the carbon slurry from the quench tank into the carbon transfer line. This condition also occurred periodically during the subsequent plant evaluation. This condition was inevitably traced to the presence of some foreign material in the quench tank, causing blockage of the quench tank discharge line, the lower valve, or the transfer line eductor. It was necessary on several occasions to drain water from the tank, remove the lower tottings, and clean out the problem material. The foreign material ranged from 1" diameter rubber balls, which entered with the granular carbon, to pieces of paper, plastic, wrenches, pens, bolts, and nuts. This problem was caused primarily by the open top quench tank being located below an opening in the floor.

The original design of the Carbon Adsorber provided for about five percent of the carbon to be removed each week from the bottom of the Adsorber and stored in the Spent Tank. Regenerated, or fresh carbon, from the Quench Tank was to be added to the top of the Adsorber. Approximately once a month the spent carbon was to be thermally regenerated in the fluid bed reactor then stored in the Quench Tank until needed to refill the Adsorber. Several problems developed during the start-up period which led to changes in the design operation.

[1] The four effluent screens at the top of the Adsorber each consisted of a cylindrical backing screen covered with a 60 mesh twill woven wire screen. They were flanged on one end and inserted into the Adsorber via four nozzles. These screens gradually plugged with carbon fines and biogrowth, and had to be cleaned every two weeks, a four-hour job. To facilitate cleaning, the rigid pipe on the discharge manifold was replaced with pressure hose with quick disconnects mating with quick disconnects welded to the flanged end of the screens. This reduced cleaning time to about one hour. Finally, the screens were replaced with wedge wire well points with 0.010 inch openings. These screens require cleaning every four to six months.

[2] In April of 1973, after approximately six months of operation, pressure drop began to be evident throughout the entire column rather than only across the effluent screens. High levels of iron and turbidity in the plant effluent were also observed periodically, in addition to the presence of hydrogen sulfide in the carbon column effluent. These symptoms appeared to result from bacteriological growth and anaerobic activity within the carbon column. The first attempt to correct this problem was to employ recirculation through the column back

to the adsorber feed tank. This tended to reduce, but not eliminate, the anaerobic frequency; however, the pressure drop increased. The addition of chlorine at a rate of 8 to 14 mg/l to the Adsorber Feed Tank was then initiated. This reduced the rate at which the pressure drop increased, especially the pressure differential across the lower third of the Adsorber. The dissolved oxygen content of adsorber effluent was still very low, at 0 to 3 mg/l, and the presence of H_2S in the adsorber effluent persisted. A compressed air line and rotometer were installed, and air was introduced into the Adsorber with the feed water at a rate of about 1.5 SCFM. The dissolved oxygen content of the adsorber effluent increased to about 8 to 10 mg/l and the anaerobic condition was eliminated.

The Standard Operating Procedure was modified to require continuous addition of chlorine at about 8 to 14 mg/l to the Adsorber Feed Tank, and air at a rate of about 2 SCFM to the base of the Adsorber. In addition, once a week about half a blow case of carbon was moved from the Adsorber to the Spent Tank, the Adsorber flushed for two hours with water and an air scour, then reloaded with carbon which had been previously washed in the Quench Tank. This procedure has produced good Adsorber operation. The BOD of the plant effluent has consistently averaged less than 5 mg/l each month; thus it has not been necessary to thermally regenerate the carbon.

(h) Chlorination

It was noted near the end of the start-up period that sludge was accumulating in the bottom of the Chlorine Contact Tank. Two nozzles 180° apart were welded on the side of the tank, flush with the bottom, and equipped with valves so the tank could be washed as needed. The chlorinated water circulating pump used to educt chlorine into the water failed. A new pump was supplied by the manufacturer. During the interval when the pump was out of service city water was used as the motive force to educt the chlorine.

(i) Emergency Generator

Power failures occurred frequently during the last four months of the start-up period. On each failure the emergency generator started and supplied power to continue wastewater treatment operations automatically without requiring operator attention. This generator starts fifteen seconds after a power failure. All wastewater processing equipment and controllers then start in sequence. The generator continues to operate until the power has been restored then shuts down automatically.

The Incinerator and related equipment do not restart automatically after a power failure, manual restart was used to assure safety.

(j) Miscellaneous

The temperature in the building would frequently exceed 38^{o} C on warm days in the early summer. A two-speed ventilating fan was installed in the roof above the Incinerator. This reduced the building temperature to acceptable levels.

The original sump pump in the basement had inadequate capacity, and frequently plugged with fibrous and solid materials. A higher capacity open impeller submersible type pump was installed which performed satisfactorily.

The magnetic filter would start its programmed air and water wash cycle before the relatively slow moving automatic valve in the effluent line to the Adsorber Feed Tank had fully closed. This would send a slug of dirty water to the feed tank, and the force of the air caused splashing. A limit switch was installed on the automatic effluent valve to delay the wash until the valve was fully closed. The air-water wash also caused a surge of water to flow back up the line to the Clarifier. A fast-action check valve was installed in the line to stop this backflow.

d. Wastewater Characteristics

(1) The occupancy rate and flow data are contained in Table 4 and are plotted in Figure 15. The high average daily flows per house for October and November 1972 were due partly to use of sewage from the force main and partly to infiltration. The flows from December 1972 through May 1973 show a steadily decreasing infiltration rate as the result of intensive efforts by the developer to locate and repair breaks in the sewerage system. The flow per house for the final seven months of 1973 was 216 gallons per day.

(2) Raw sewage, surge tank, and plant effluent analysis are contained in Tables 5, 6 and 7 respectively. The BOD_5 and suspended solids values are plotted in Figures 16 and 17 The influent, typical of domestic sewage, became stronger as the infiltration decreased and the occupancy rate increased. Despite various start-up problems, the quality of plant effluent was consistently high, with all parameters well below design limits. These data are the average of daily composites.

2. Steady State Operations

By June, 1973, most of the houses in the development were occupied, influent sewage flow had stabilized at about 26,000 GPD, and a number of problems in the liquid handling process had been encountered and corrected. A 36-week period of "steady-state" operation was designated to begin on July 1, 1973.

TABLE 4

OCCUPANCY AND FLOW
WOODGATE FARMS - FREEHOLD, N. J.

	Average Daily Flow Gallons	Titles Issued During Month	Cumulative Titles Issued	Average Monthly Occupancy	Avg. Flow Per House Gal. /Day
October '72	6,928(1)	14	14	7	990(1)
November '72	21,503(1)	12	26	20	1,075(1)
December '72	18,192(1)	21	47	37	492(1)
January '73	16,331	10	57	52	314
February '73	24,307	10	67	62	392
March '73	24,615	11	78	73	337
April '73	26,836	8	86	82	327
May '73	27,429	14	100	93	295
June '73	26,168	13	113	107	245
July '73	25,344	7	120	117	217
August '73	25,917	1	121	121	214
September '73	26,816	3	124	123	218
October '73	24,324	0	124	124	196
November '73	26,702	0	124	124	215
December '73	26,053	2	126	125	208

Note: (1) Include sewage from another community pumped in for start-up testing.

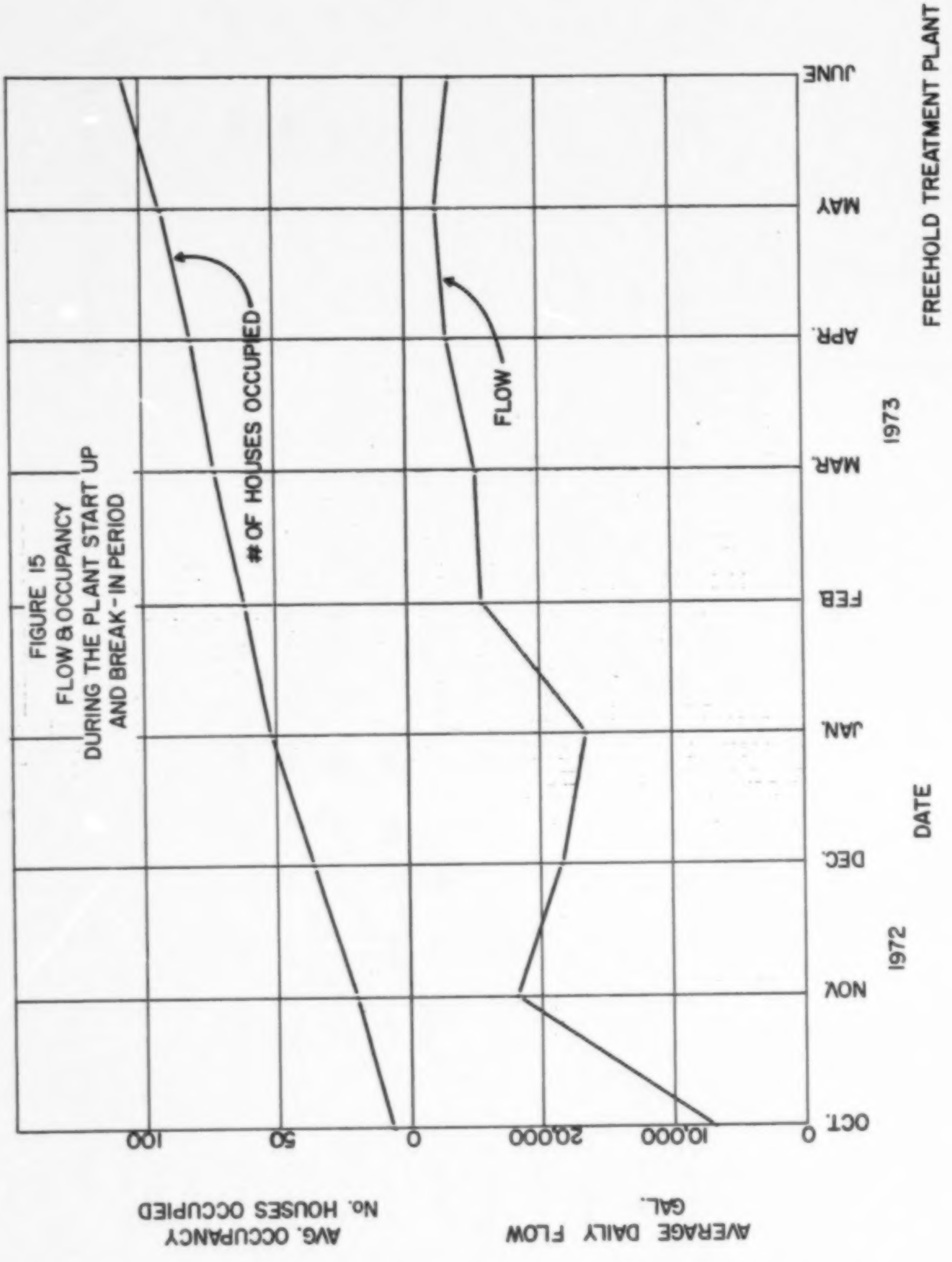


FIGURE 15
FLOW & OCCUPANCY
DURING THE PLANT START UP
AND BREAK-IN PERIOD

TABLE 5

RAW SEWAGE CHARACTERISTICS

Month	pH	SS	TOC	BOD_5	COD	TOD	P	TKN	NH_3-N	NO_2 + NO_3	Acidity	Alk.	Total Solids	Total Solids % Ash
October '72	7.0	266	14	36	93	-	-	-	-	-	28	34	-	-
November '72	6.9	119	-	-	-	-	-	-	-	-	12	21	-	-
December '72	7.2	185	-	-	152	-	4.3	25.6	28.8	-	-	-	-	-
January '73	7.5	147	-	52	162	-	6.4	35.7	28.0	4.5	48	-	-	-
February '73	7.3	212	46	-	242	272	-	-	-	-	-	-	-	-
March '73	7.3	165	-	165	356	-	-	-	-	-	-	-	-	-
April '73	7.0	280	-	132	345	-	6.5	-	-	-	-	-	504	34
May '73	7.5	208	-	186	333	-	4.8	-	-	-	-	-	434	52
June '73	7.6	191	-	83	403	-	6.3	-	-	-	-	-	493	43

NOTE: All units mg/l except as noted.

FREEHOLD TREATMENT PLANT

TABLE 6

SURGE TANK CHARACTERISTICS

Month	pH	SS	TOC	BOD_5	COD	TOD	P	TKN	NH_3-N	NO_2 + NO_3	Acidity	Alk.	Total Solids	Total Solids % Ash
October '72	7.3	123	17	22	80	-	1.04	14.6	10.0	0.21	21.6	93	-	-
November '72	-	140	-	-	-	-	1.37	17.7	14.8	2.1	17.3	106	-	-
December '72	7.2	120	-	-	151	-	4.44	24.2	14.8	11.4	-	-	-	-
January '73	7.5	147	40	86	299	-	2.55	29.4	16.5	6.3	49.7	173	-	-
February '73	7.2	153	38	98	216	284	1.53	38.5	16.5	2.1	43.2	165	-	-
March '73	7.1	181	40	195	252	185	3.03	19.0	12.4	0.6	-	-	-	-
April '73	7.0	240	39	143	276	324	3.43	47.0	25.5	1.7	-	175	816	46
May '73	7.5	200	38	154	332	283	4.57	50.0	22.2	2.8	-	150	442	51
June '73	7.5	230	53	77	391	354	7.01	55.0	27.2	2.6	-	158	489	41

NOTE: All units mg/l except as noted.

FREEHOLD TREATMENT PLANT

TABLE 7

PLANT EFFLUENT CHARACTERISTICS

Month	pH	Fe	Free Res. Cl_2	SS	TOC	BOD	COD	TOD	P	TKN	NH_3-N	NO_2 + NO_3	Total Solids	Total Solids %Ash	Alk.	Col.	Avg. Daily Flow Gal.
Oct. '72	6.8	-	0.4	3.2	4	1	9	-	.26	7.9	3.9	.03	-	-	41	0	6,928
Nov. '72	6.6	-	0.5	1.8	-	3	1	-	-	-	-	-	-	-	42	-	21,503
Dec. '72	6.6	-	0.9	1.3	-	-	12	-	.17	17.0	12.4	8.1	-	-	-	-	18,192
Jan. '73	7.1	3.2	0.9	6.1	11	1	15	52	.95	20.4	11.1	2.5	-	-	97	-	16,331
Feb. '73	7.2	0.7	1.6	2.4	7	5	44	33	.19	20.9	12.2	1.0	-	-	134	0	24,307
Mar. '73	6.9	1.4	0.9	4.1	7	1	17	22	.65	19.0	10.8	0.5	-	-	-	1	24,615
Apr. '73	6.7	1.4	0.7	4.1	6	1	20	49	.46	18.5	15.8	0.9	335	24	91	0	26,836
May '73	7.0	1.4	2.2	4.7	5	4	26	53	.26	40.0	16.5	1.7	381	43	96	0	27,429
June '73	7.0	0.9	1.5	4.0	6	1	33	49	.23	38.0	21.4	1.6	348	30	117	0	26,168

NOTE: All units mg/l except as noted.

FREEHOLD TREATMENT PLANT

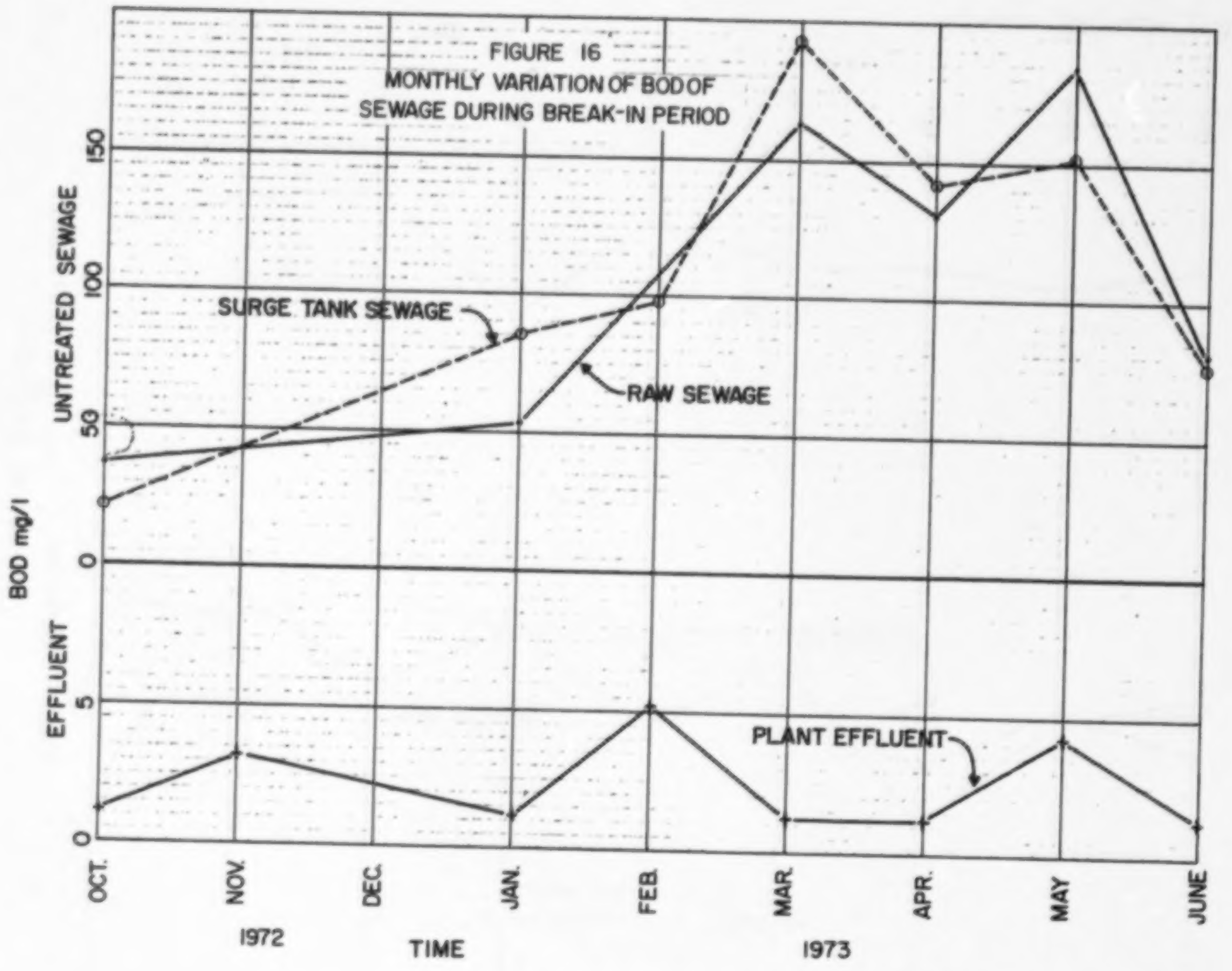


FIGURE 16
MONTHLY VARIATION OF BOD OF
SEWAGE DURING BREAK-IN PERIOD

FREEHOLD TREATMENT PLANT

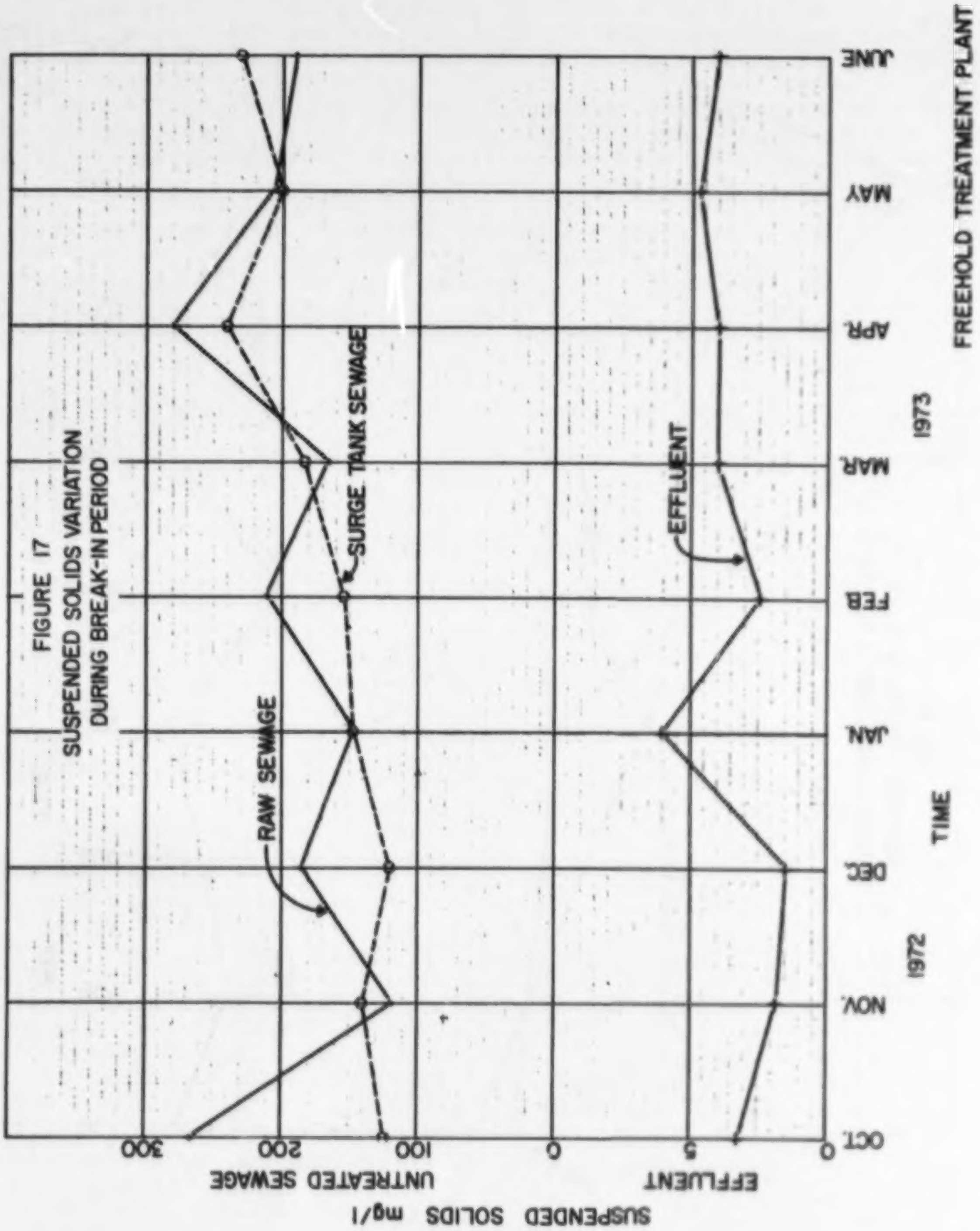


FIGURE 17
SUSPENDED SOLIDS VARIATION
DURING BREAK-IN PERIOD

FREEHOLD TREATMENT PLANT

A schedule of normal sampling and analysis was established in accordance with Table 8. A schedule of intensive evaluation was also specified as shown in Table 9. The intensive tests were to be conducted during eight one-week periods, in addition to the normal evaluation.

Several changes in these schedules were made during the course of the test period. Some tests were added or increased in frequency, while others were dropped, based on evaluation of the results collected.

a. Normal Evaluation

The analytical schedule for the normal evaluation included (1) a rather complete set of samples and analyses for one day each week, (2) somewhat reduced coverage for two additional days, and (3) minimal analyses on the remaining four days. A rotating sampling plan was used in order to cover all days of the week during the 36-week duration of the evaluation. Results were reported weekly, using the analysis report form typified by Figure 18. Analytical and operating data were also summarized and reported on a monthly basis.

(1) Raw Sewage Characteristics

(a) Flow

As noted above, the variable and abnormally high average raw sewage flow experienced during the start-up and break-in period had diminished by June, 1973. This stabilization in influent flow was due primarily to correction of the infiltration problems encountered during construction of the houses in the development. Occupancy and flow data for the steady-state period are shown in Table 10; the daily flow per home during this nine-month period averaged 207 gallons, and ranged from 173 to 226 on a monthly basis. These data are shown graphically in Figure 19, including the latter six months of the break-in period from January through June, 1973.

A frequency distribution of the 274 daily effluent flows occurring during the steady state phase is shown in Figure 20. These flows are the daily differences read from the plant effluent totalizing meter at 8:00 a.m. each day. Approximately 90% of the values fell within the range of 18,000 to 33,000 gallons per day. It should be noted, however, that these flows are not corrected for changes in surge tank level, nor do they reflect the effect of the frequent failure and erratic operation of the lift station. Limited study of these factors showed that the extremes in effluent flow, both high and low, were always related either to a lift station failure or to some other condition causing a substantial change in surge tank level. These changes in turn required frequent adjustment

TABLE 8

ANALYTICAL & SAMPLING SCHEDULE DURING NORMAL EVALUATION

Sample Point	Analyses & Frequency
Hydra Sieve Feed (Raw Sewage)	pH, Alkalinity, Suspended Solids, BOD, TOC or COD, Total P, Total Nitrogen, Ammonia Nitrogen, Coliform 3 times per week on 24-hour composites
Hydra Sieve Effluent (Holding Tank)	Same as above + Soluble TOC or COD + Soluble P Once per week on 24-hour composite
Clarifier Effluent	pH, Suspended Solids, Total P, Coliform Once per week on 24-hour composite
Ferrofilter Effluent	pH, Alkalinity, Suspended Solids, Total COD or TOC, Soluble TOC or COD, Total P, Soluble P, Total N, Ammonia N, Coliform Once per week on 24-hour composite Dissolved O_2 once per day (grab sample)
Carbon Column Effluent	pH, Total & Soluble TOC or COD, Fe Once per week on 24-hour composite D.O. once per day (grab sample)
Plant Effluent	Same as Hydra Sieve Effluent + Fe + Cl_2 residual 3 times per week on 24-hour composite
Hydra Sieve Solids Clarifier Sludge Dewatered Sludge	Total Solids, Volatile Solids Once per week

GENERAL: ALL DATA ON LIQUID AND SLUDGE FLOW, FUEL CONSUMPTION, ELECTRICITY USE AND MANPOWER REQUIREMENT TO ALLOW FOR ECONOMIC ANALYSIS.

FREEHOLD TREATMENT PLANT

TABLE 9

ANALYTICAL & SAMPLING SCHEDULE IN ADDITION TO NORMAL EVALUATION DURING INTENSIVE EVALUATION

Sample Point	Frequency	Analyses
Hydra Sieve Feed	2 days/week	pH, Alkalinity, Suspended Solids, BOD, TOC or COD, Total P, Total N, Ammonia N, Coliform. Each on 2-6 hour and one 12-hour composite
Hydra Sieve Effluent	2 days/week	Same as above + Soluble TOC or COD and Soluble P. Each on 2-6 hour composites and one 12-hour composite
Clarifier Effluent	2 days/week	pH, Suspended Solids, Total P, Coliform Each on 2-6 hour composites, and one 12 hour composite
Ferrofilter Effluent	2 days/week	pH, Alkalinity, Suspended Solids, Total & Soluble COD or TOC, Total P, Soluble P, Total N, Ammonia N, Coliform Each on 2-6 hour composites and one 12-hour composite
Carbon Column Effluent	2 days/week	pH, Total & Soluble TOC or COD, Fe Each on 2-6 hour composites and one 12-hour composite
Plant Effluent	2 days/week	Same as Hydra Sieve Effluent + Fe + Cl_2 residual Each on 2-6 hour composites and one 12-hour composite
Solids Handling Loop	Total week	Complete Solids Balance & Sludge Volume Accounting

FREEHOLD TREATMENT PLANT

FIGURE 18

FREEHOLD

WEEKLY ANALYSIS REPORT

12/14/73

	Sludge		
	Hydrasieve	Clarifier	Hold Tank
Total Solids %	17.4	4.7	4.4
Volatile Solids %	94.1	48.0	48.6

		pH	SS	Fe	DO	Res. Cl$_2$ Free	Res. Cl$_2$ Total	COLI. MPN	Tot. TOC	Sol. TOC	BOD	COD	TOD	TKN	NO$_2$ + NO$_3$-N	NH$_3$-N	Tot. P	Sol. P	ALK.	Chloride	Tot. Solids	Vol. Solids
	R	7.8	590					2.3×10^6	140		384			72.6		36.9	14.6		200			
SUN.	ST																					
Dec.	CE																					
9	FFE				10																	
	CCE				8																	
23,386 gal.	PE	7.3	0.9	0.2		3.0	7.0	<3	4	4	0			28.0		24.6	.047	.047	128			
	R	8.7	108																			
MON.	ST																					
Dec.	CE		10.0																			
10	FFE		10.6		9																	
	CCE				9																	
28,122 gal.	PE	6.8	2.5	0.3		0.8	1.9															
	R	8.2	328					9.3×10^5	96		240			69.0		37.1	9.25		206			
TUES.	ST	7.7																				
Dec.	CE	7.1																				
11	FFE	7.1																				
	CCE	6.7																				
24,504 gal.	PE	6.8	1.9	0.5		0.9	2.0	<3	6	4	0			20.6		14.8	.056	.038	124			
	R	8.2																				
WED.	ST																					
Dec.	CE		8.3																			
12	FFE		4.0		10																	
	CCE				9																	
17,776 gal.	PE	6.9	1.9	0.3	9	1.0	2.0															
	R																					
THURS.	ST																					
Dec.	CE																					
13	FFE																					
	CCE																					
23,499 gal.	PE					0.8	1.2															
	R	6.9	622					2.3×10^5	152		228	1024	660	72.6		25.4	18.0		158	158	880	405
FRI.	ST	7.0	400					7.5×10^5	144	20	132	552	600	54.4		31.9	17.0	12.0	194			
Dec.	CE	7.2	11					7.5×10^4									.093					
14	FFE	7.0	17		9			9×10^4	28	16	12	55.1	126	30.6	3.3	24.6	.074	.047	168			
	CCE	6.8			9				12	7												
24,551 gal.	PE	7.0	3	1.5		3.0	8.0	23	10	8	2	63	84	28.0	1.5	24.6	.056	.047	130	180	533	42
	R	7.7	152																			
SAT.	ST																					
Dec.	CE																					
15	FFE				10																	
	CCE				9																	
23,666 gal.	PE	7.0	4.3	1.1		1.0	2.5															

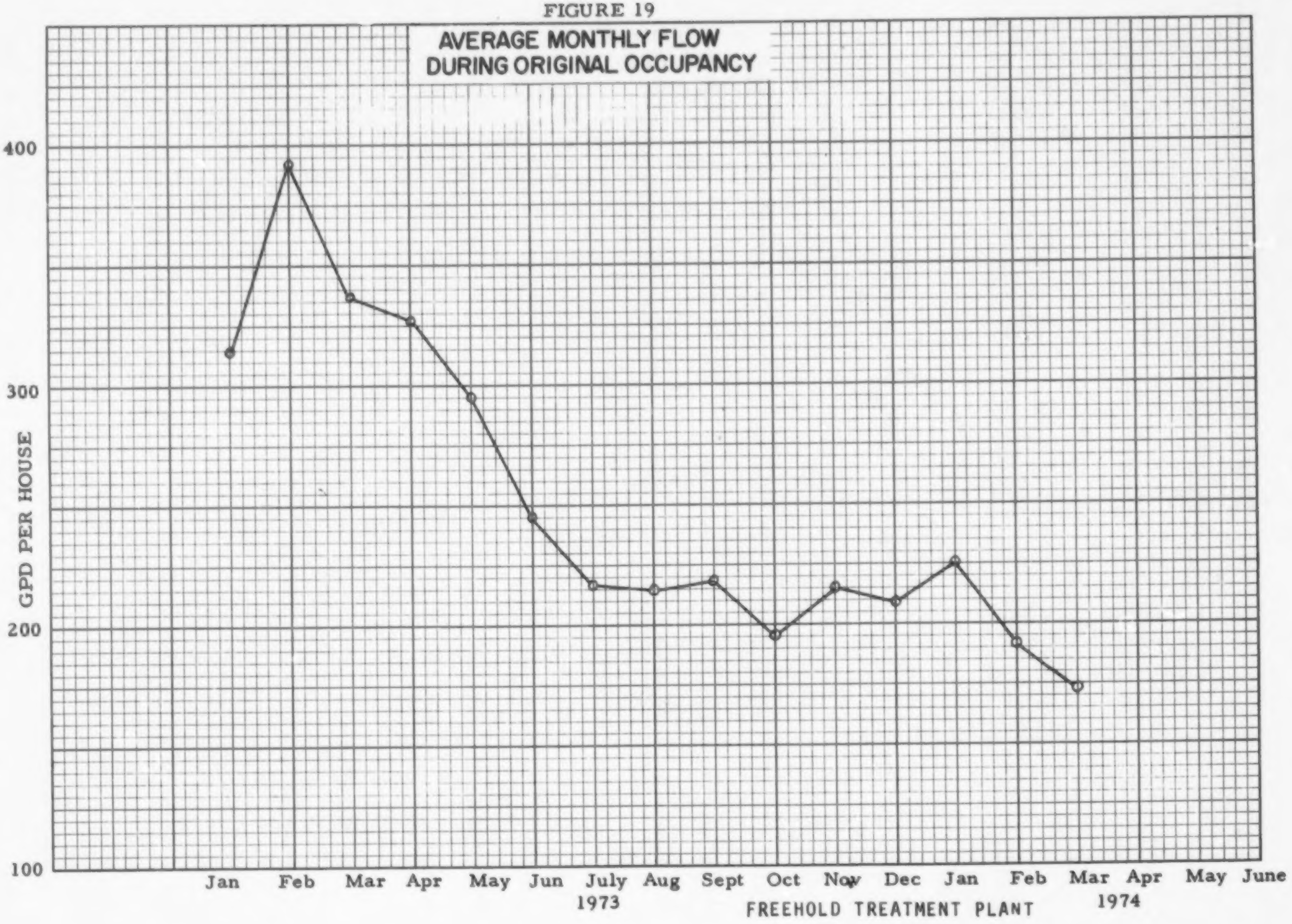
TABLE 10

OCCUPANCY AND FLOW

Month	Average Daily Flow Gallons	Cumulative Titles Issued	Average Monthly Occupancy	Average Flow Per House Gallons/Day
1973				
July	25,344	120	117	217
August	25,917	121	121	214
September	26,816	124	123	218
October	24,324	124	124	196
November	26,702	124	124	215
December	26,053	126	125	208
1974				
January	28,490	126	126	226
February	24,134	126	126	192
March	21,737	126	126	173

FREEHOLD TREATMENT PLANT

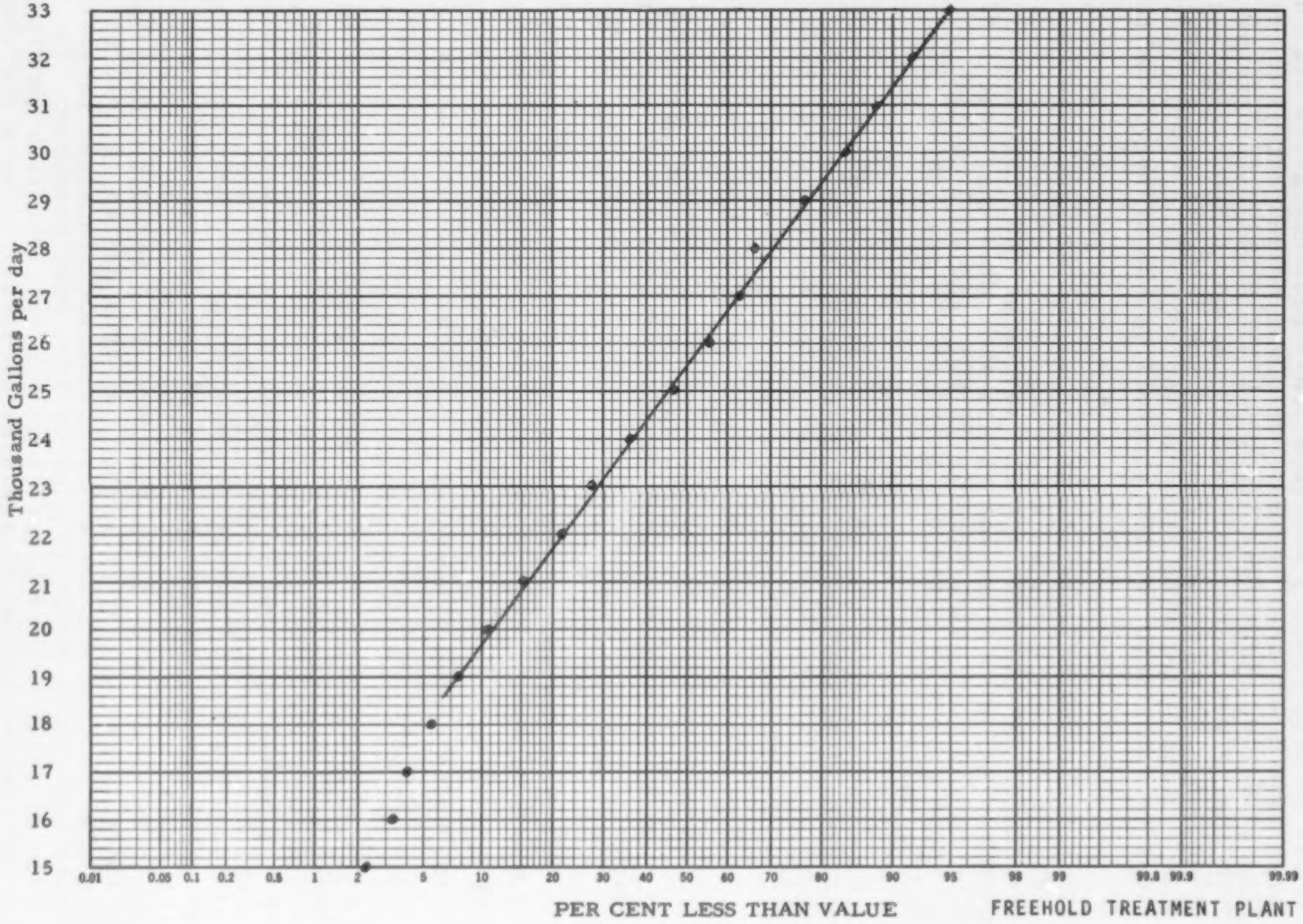
FIGURE 19



AVERAGE MONTHLY FLOW
DURING ORIGINAL OCCUPANCY

FIGURE 20

DAILY EFFLUENT FLOW JULY 1973-MARCH 1974
FREQUENCY DISTRIBUTION



in the process flow to maintain continuous operation without a surge tank overflow. The most frequent normal operating condition affecting surge tank level was the weekly transfer of carbon from the carbon column followed by carbon column backwash and reloading. This sequence required the internal use of substantial process water with accompanying adjustments in operating flows to compensate for the abnormal volume of recycle water.

Correction of the recorded effluent flows for these effects, to obtain daily influent flow, would undoubtedly have produced a narrower range of values. This correction was not made for the entire period, but study of several minimum and maximum values indicated that true influent flow ranged from approximately 17,000 to 34,000 gallons per day, or roughly plus or minus one third of the mean daily flow of 25,315 gallons.

A 24-hour influent flow profile for the plant was developed by averaging 10 selected days of surge tank level and effluent flow data. The smoothed curve is shown in Figure 21.

(b) Raw and Surge Tank Sewage Quality

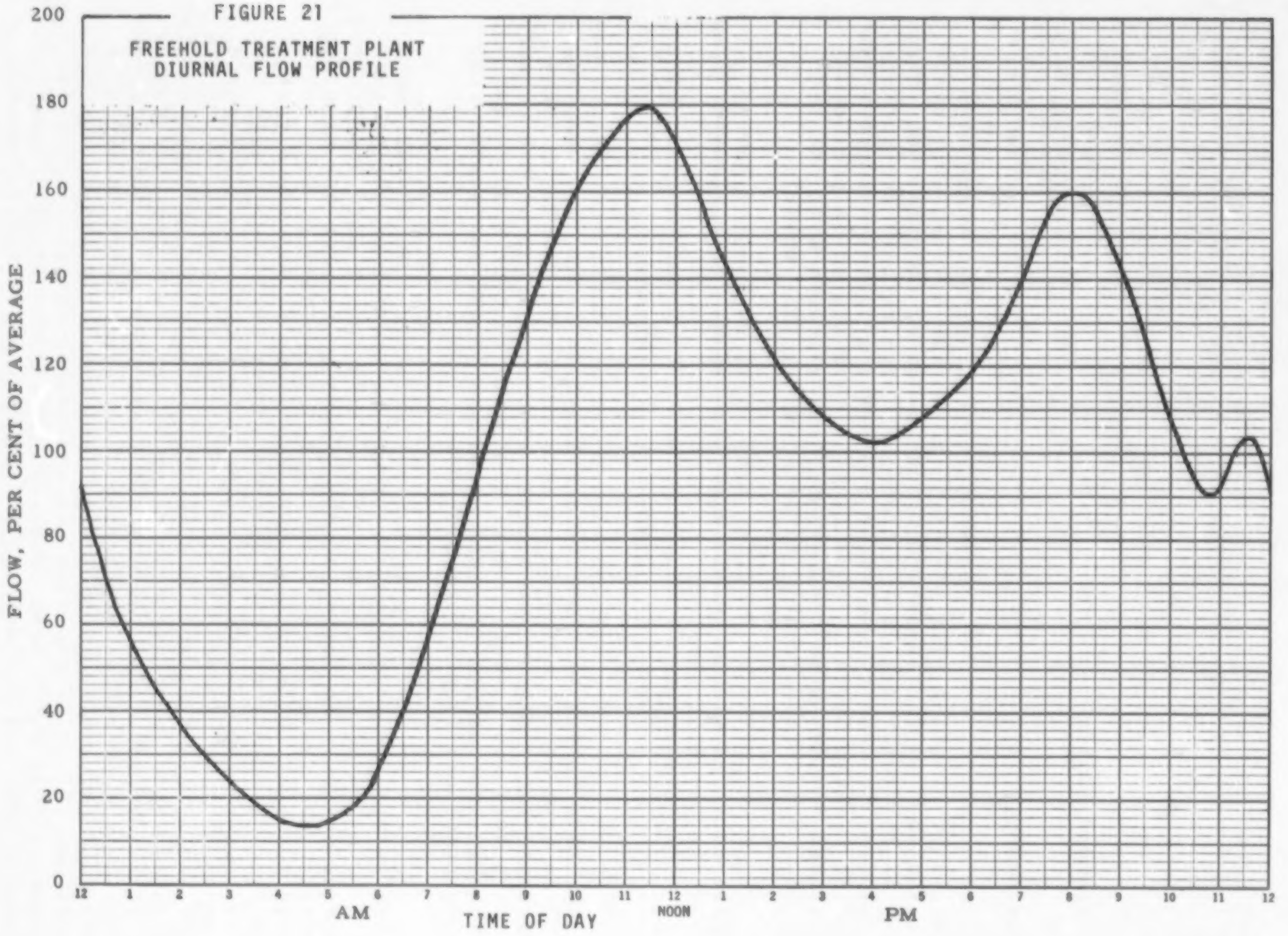
Raw sewage and surge tank averages for four major parameters determined during this period are shown below:

	Raw	Surge Tank
Suspended Solids	242	182
BOD	207	163
Total Hydrolyzable Phosphorus	10.8	8.3
Total Kjeldahl Nitrogen	44.7	53.1

Raw sewage sampling was subject to the usual difficulties in obtaining representative composites, and was discontinued in December, 1973, with a corresponding increase in the number of Surge Tank samples. The two sets of determinations are, therefore, not strictly suited for direct comparison, although average values for suspended solids, BOD, and phosphorus did show similar reductions (25, 21 and 23% respectively) from the raw to surge tank samples. The suspended solids reduction is in agreement with calculated removals, based on the weight of primary screen solids collected during a period of several days. The same pattern was not evident, however, in other parameters where comparable reductions might have been expected. The absence of other correlations is undoubtedly due to a combination of the limited number of analyses performed, sample variability, and analytical precision.

FIGURE 21

FREEHOLD TREATMENT PLANT
DIURNAL FLOW PROFILE



Frequency distributions of raw sewage and surge tank analyses for suspended solids, BOD, and phosphorus are shown in Figures 22,23 and 24. As seen from these graphs the median values and 80% occurence for these properties in the raw sewage were approximately:

	Median	80% Range
Suspended Solids	200	120-400
BOD	190	100-350
Phosphorus	11	4-18

Nitrogen determinations showed low levels of nitrites plus nitrates in the raw sewage, averaging less than 2 mg/l. Ammonia nitrogen averaged somewhat less than 50% of the total Kjeldahl nitrogen. These results are not unexpected, in view of the close proximity of the treatment plant to the sewage source. There is some evidence of increased NH_3 nitrogen in the surge tank as would be expected with the 8 to 12 hour detention normally experienced in this vessel.

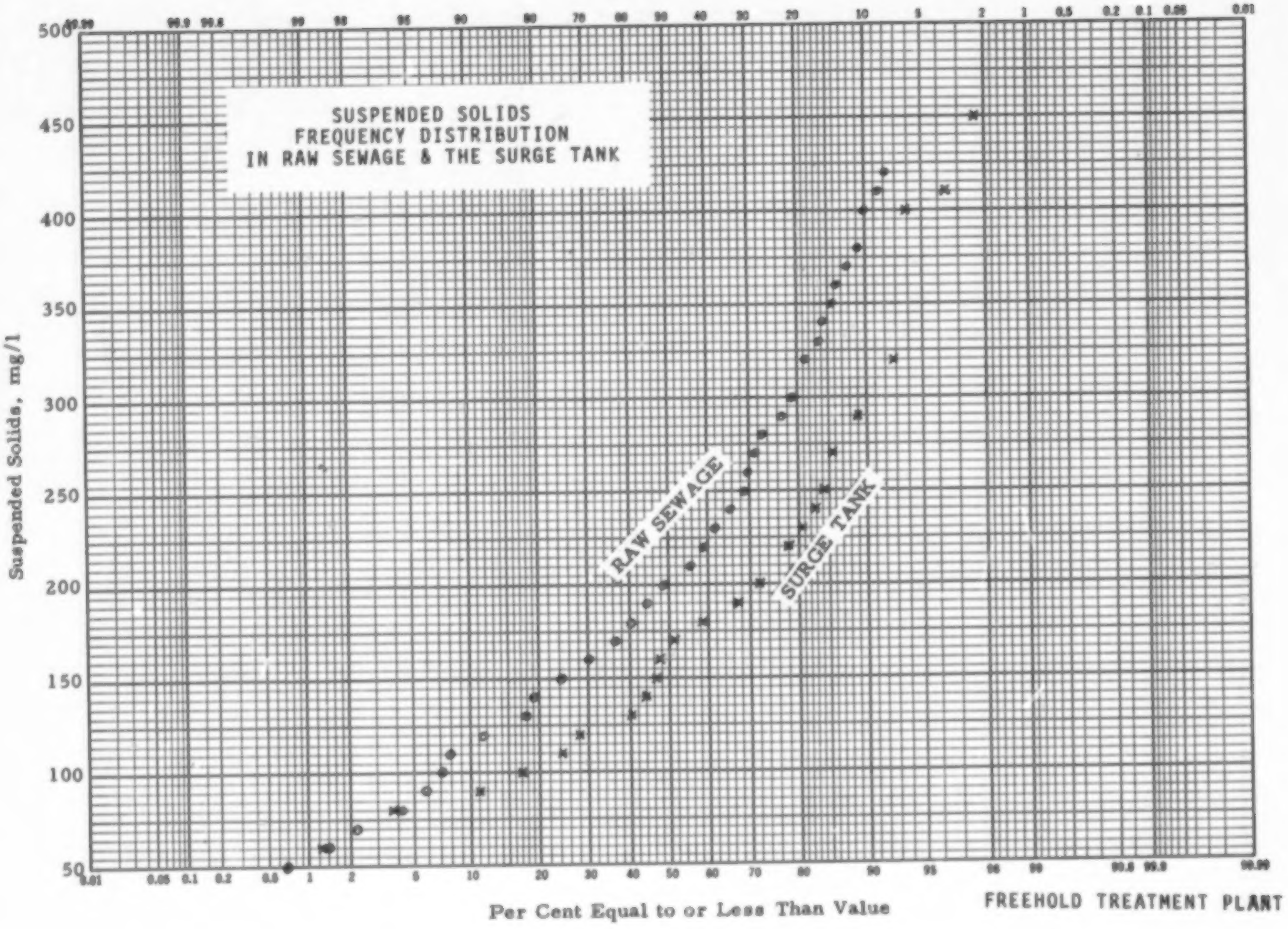
Other properties of the raw sewage appeared to be within normal expected ranges for wastewater from a purely domestic source.

(2) Liquid Treatment Performance

Average values for all analytical determinations made during this period are shown in Table 11. Sampling points were at the effluent of each vessel or location listed with the exception of raw sewage, which was taken at the primary acreen head box. Samples were generally 24 hour composites, although grab composites were occasionally necessary when sampling pumps or lines became plugged.

As previously indicated, some changes were made in the sampling schedule during the course of the steady state period of evaluation. These changes were made to ease the analytical load in areas where differences between two sampling points were small or where continued determinations were not felt to be of sufficient value. For example: nitrite-nitrate analyses were discontinued because values were always very low; raw sewage samples were discontinued in December, 1973 because the surge tank sample was felt to be more representative of the loading on the treatment processes. Ferrofilter effluent samples were no longer meaningful when the filter was taken out of service in January 1974. Because of these changes certain analyses have been combined in evaluating plant performance in order to increase the number of determinations available and thus minimize the effect of sampling and analytical variability. Although approximately 3,300 major analyses (excluding pH, dissolved oxygen, iron and residual chlorine) were performed during the

FIGURE 22



FREEHOLD TREATMENT PLANT

FIGURE 23

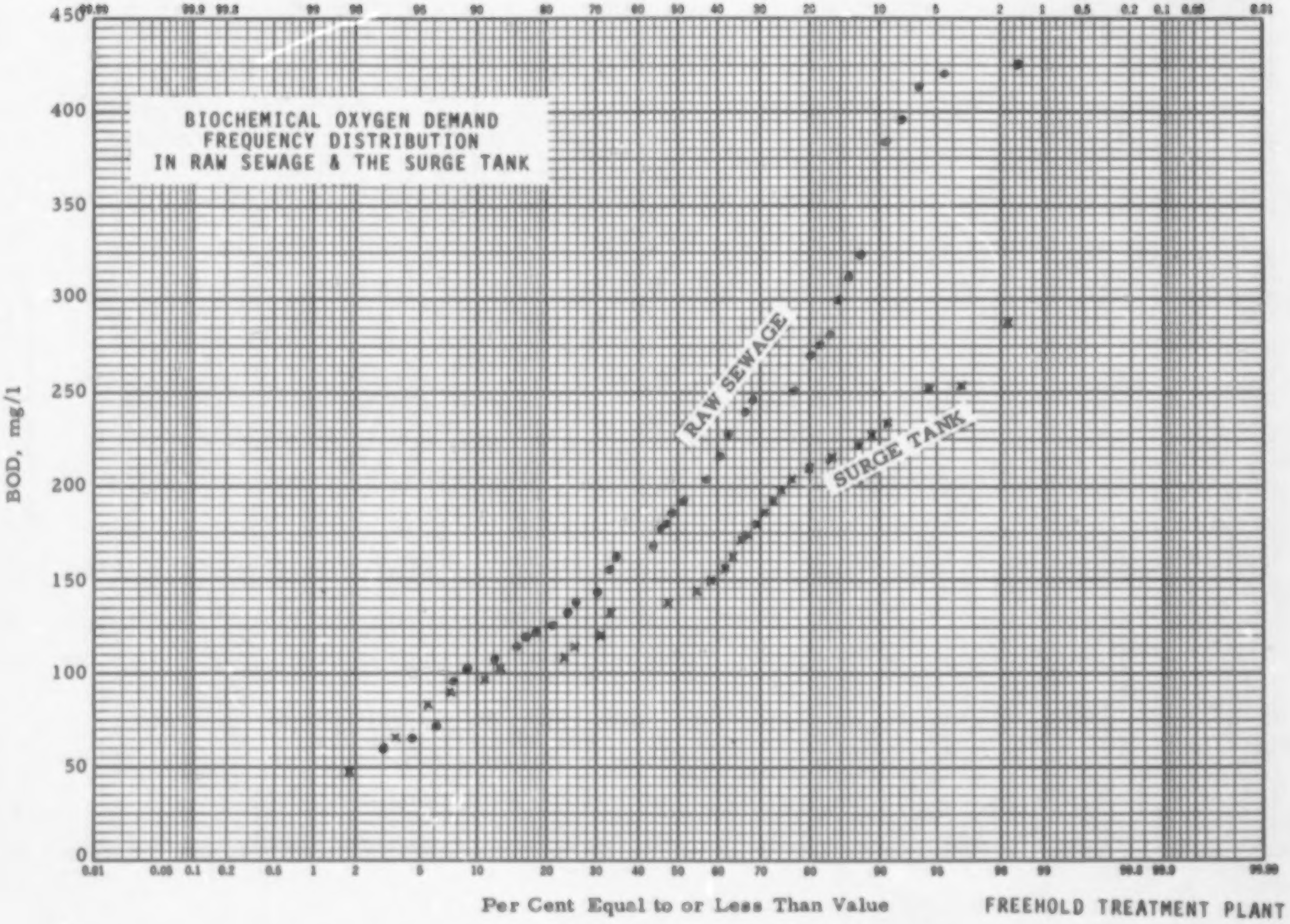

BIOCHEMICAL OXYGEN DEMAND
FREQUENCY DISTRIBUTION
IN RAW SEWAGE & THE SURGE TANK
RAW SEWAGE
SURGE TANK
BOD, mg/l
450
400
350
300
250
200
150
100
50
0
99.99
99.9
99.8
99
98
95
90
80
70
60
50
40
30
20
10
5
2
1
0.5
0.2
0.1
0.05
0.01
Per Cent Equal to or Less Than Value
FREEHOLD TREATMENT PLANT

FIGURE 24

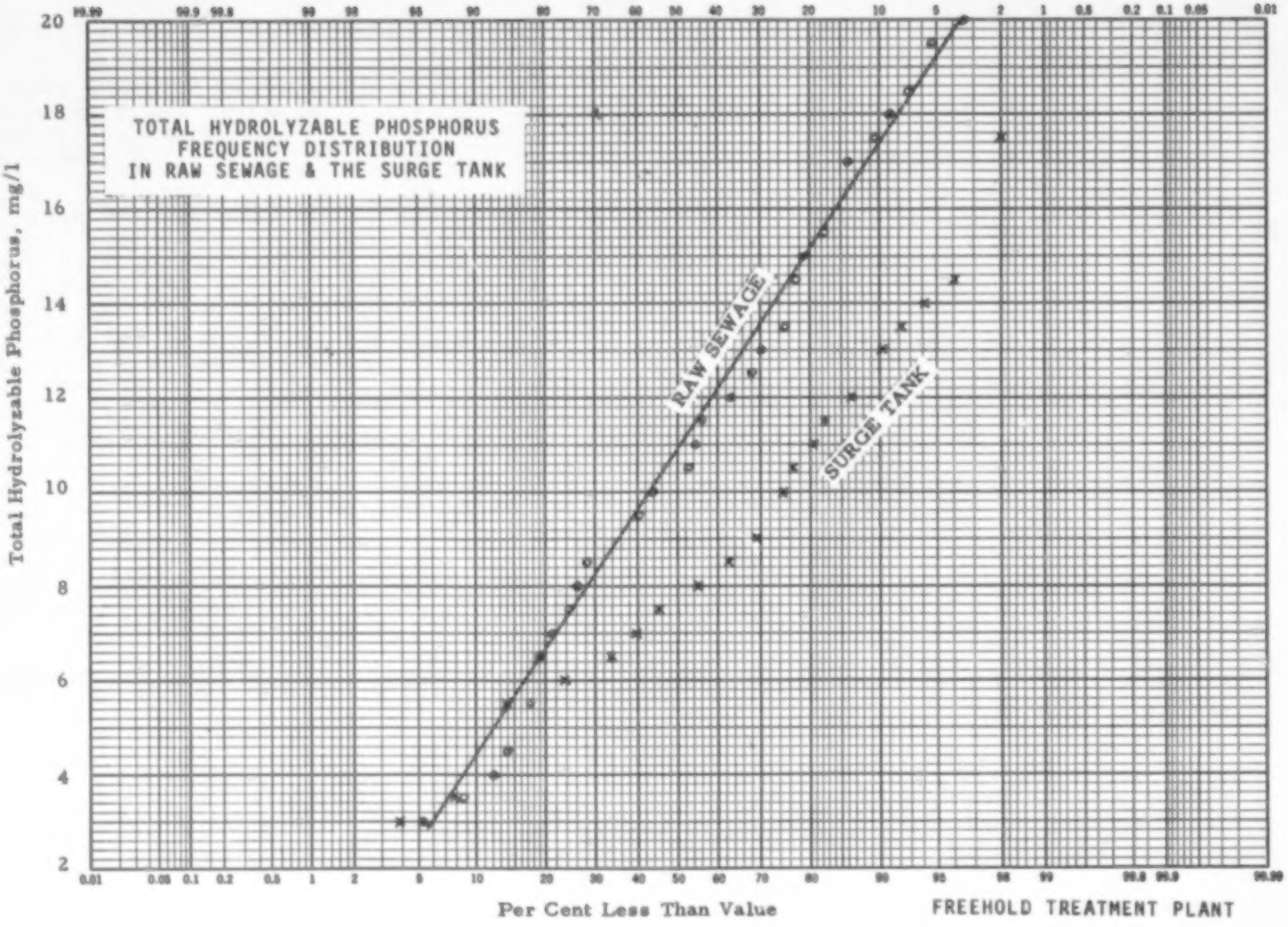

TOTAL HYDROLYZABLE PHOSPHORUS
FREQUENCY DISTRIBUTION
IN RAW SEWAGE & THE SURGE TANK
RAW SEWAGE
SURGE TANK
Total Hydrolyzable Phosphorus, mg/l
20
18
16
14
12
10
8
6
4
2
99.99
99.9
99.8
99
98
95
90
80
70
60
50
40
30
20
10
5
2
1
0.5
0.2
0.1
0.05
0.01
Per Cent Less Than Value
FREEHOLD TREATMENT PLANT

TABLE 11

LIQUID TREATMENT PERFORMANCE AVERAGE VALUES

Analysis, mg/l	Raw Sewage	Surge Tank	Clarifier	Ferrofilter	Carbon Column	Plant Effluent
Suspended Solids	242	182	12.2	10.0	-	2.9
BOD	207	163	-	35.1	-	1.4
TOC	72	74	-	20.1	5.7	5.6
Soluble TOC	-	27	-	10.3	4.9	4.1
TOD	460	433	-	143	-	72
COD	540	463	-	103	-	25
TKN	44.7	53.1	-	40.2	-	28.8
NO_2 + NO_3-Nitrogen	1.8	-	-	1.4	-	1.4
NH_3-Nitrogen	19.3	28.0	-	24.3	-	17.3
Phosphorus	10.8	8.3	0.38	0.31	-	0.16
Soluble Phosphorus	-	5.6	-	0.26	-	0.19
Alkalinity	202	205	-	151	-	110
Chloride	78	105	-	-	-	233
Total Solids	628	614	-	-	-	482
Volatile Solids	324	314	-	-	-	111
Total Coliform (MPN per 100 ml)	1.1×10^6	2.2×10^6	1.4×10^5	6.1×10^4	-	<3.5
Iron	-	-	-	-	0.9	0.7
pH (units)	7.5	7.4	7.1	7.0	6.9	6.9
Dissolved Oxygen	-	-	-	8	8	9
Free Residual Chlorine	-	-	-	-	-	1.1
Total Residual Chlorine	-	-	-	-	-	3.4

FREEHOLD TREATMENT PLANT

nine month period, an even greater number of determinations would have added to the confidence level in evaluating overall performance and unit operation efficiencies.

Average results for the four major parameters are summarized in Table 12 where Raw Sewage and Surge Tank values have been combined for Kjeldahl nitrogen, and Clarifier Effluent and Ferrofilter Effluent have been combined for BOD and Kjeldahl nitrogen. These averages were used in calculating the efficiencies shown in Table 13. Three sets of efficiency values are shown:

(a) The individual major process step unit efficiency, that is: the percentage of pollutant removed in each step as related to the input to that operation.

(b) The cumulative removal at each process step as related to the raw sewage (or raw sewage-surge tank average in the case of Kjeldahl nitrogen). These data are also shown graphically in Figure 25.

(c) The contribution of each process step in removing a pollutant, expressed as percent of the raw influent.

Several conclusions are evident from examination of these results:

(1) Clarification (which includes chemical treatment) was the most effective step in the removal of suspended solids, phosphorus, and BOD, accounting for 70, 73 and 62% respectively of these pollutants entering with the raw sewage. Twelve percent of the Kjeldahl nitrogen was also removed by this step.

(2) Primary screening removed about a quarter of the suspended solids, phosphorus, and BOD.

(3) Carbon adsorption, although accounting for removal of only 16% of the original BOD, and 2 and 3% of phosphorus and suspended solids, was quite effective as a unit operation with efficiencies of 96, 46 and 71% respectively. Twenty-five percent of Kjeldahl nitrogen was also removed by this operation.

(4) Filtration was a relatively ineffective operation, removing only one percent of the suspended solids and phosphorus, with unit efficiency of 18%.

(5) Overall removals for the total system were 99% for suspended solids, phosphorus, and BOD. Kjeldahl nitrogen was reduced by 37%.

TABLE 12

LIQUID TREATMENT PERFORMANCE AVERAGE ANALYSES, mg/l

	Suspended Solids	Phosphorus	BOD	Kjeldahl Nitrogen
Raw Sewage	242	10.8	207	45.9
Surge Tank	182	8.3	163	
Clarifier Effluent	12.2	.380	35.1	40.2
Ferrofilter Effluent	10.0	.313		
Plant Effluent	2.9	0.161	1.4	28.8

TABLE 13

LIQUID TREATMENT PERFORMANCE EFFICIENCIES, %

	Suspended Solids	Phosphorus	BOD	Kjeldahl Nitrogen
1. Unit Efficiency as Per Cent of Unit Input				
Primary Screening	25	23	21	-
Clarification	93	95	78	12
Filtration	18	18		
Carbon Adsorption	71	46	96	28
2. Cumulative Removal				
Primary Screening	25	23	21	-
Clarification	95	96	83	12
Filtration	96	97		
Carbon Adsorption	99	99	99	37
3. Unit Efficiency as Per Cent of Raw Influent				
Primary Screening	25	23	21	-
Clarification	70	73	62	12
Filtration	1	1		
Carbon Adsorption	3	2	16	25
Residual	1	1	1	63

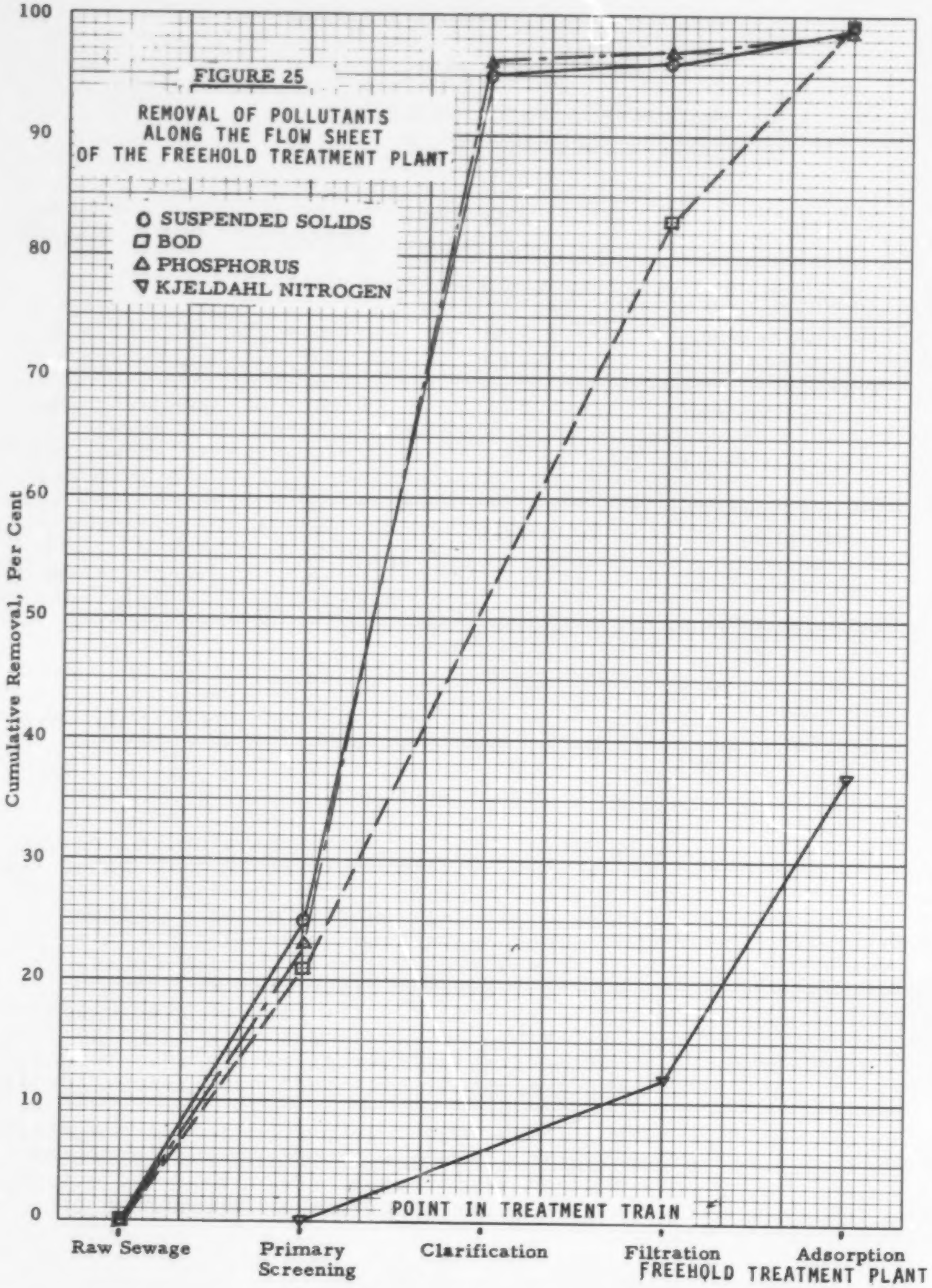


FIGURE 25

REMOVAL OF POLLUTANTS ALONG THE FLOW SHEET OF THE FREEHOLD TREATMENT PLANT

Other interesting observations of the average results show the following:

(6) A threefold increase in chloride content, as a result of the $FeCl_3$ addition.

(7) Reduction of alkalinity by 46%.

(8) Reduction of total solids by 23% despite the increase in chlorides.

(9) Ninety-four percent reduction of coliform bacteria through the filtration step. Interestingly, the unit efficiency of the filter for coliform removal was 56% compared to only 18% for suspended solids and phosphorus.

Figures 26, 27, 28, 29 and 30 show suspended solids, total phosphorus, BOD, Kjeldahl nitrogen, and NH_3-nitrogen plotted on time scales. Suspended solids is shown on a weekly basis. For this graph, raw sewage results were averaged with surge tank data, the latter increased by 25% to correct for solids removed by the primary screen. Clarifier effluent and Ferrofilter effluent data were also combined since differences between these sample points were small. The other parameters are plotted from monthly averages, again combining certain data where appropriate.

Frequency distributions for Clarifier, Ferrofilter, and plant effluent suspended solids are shown in Figure 31 and 32. The Clarifier-Ferrofilter distributions clearly show the small difference between these sample points as indicated previously by the average data. A rather sharp change in slope of the clarifier suspended solids distribution curve is evident at about the 12 to 15 mg/l suspended solids level. This break is indicative of the nature of clarification performance following chemical treatment and flocculation. The flat portion of the curve represents normal operation with a high level of suspended solids removal efficiency. The steep section shows evidence of some upset condition with sharply reduced separation efficiency. The Ferrofilter effluent show substantially lower suspended solids in this region of the frequency distribution. This difference is due primarily to the effectiveness of the turbidimetric control system following the clarification step which automatically recycled unacceptable effluent rather than to any increased Ferrofilter performance at high suspended solids levels.

The plant effluent suspended solids distribution shows a median value of 1.9 mg/l with 90% of the determination less than 6 mg/l and 96% less than 9 mg/l. This reflects the filtration achieved on the carbon column as the clarifier effluent median was 9 mg/l.

FIGURE 26

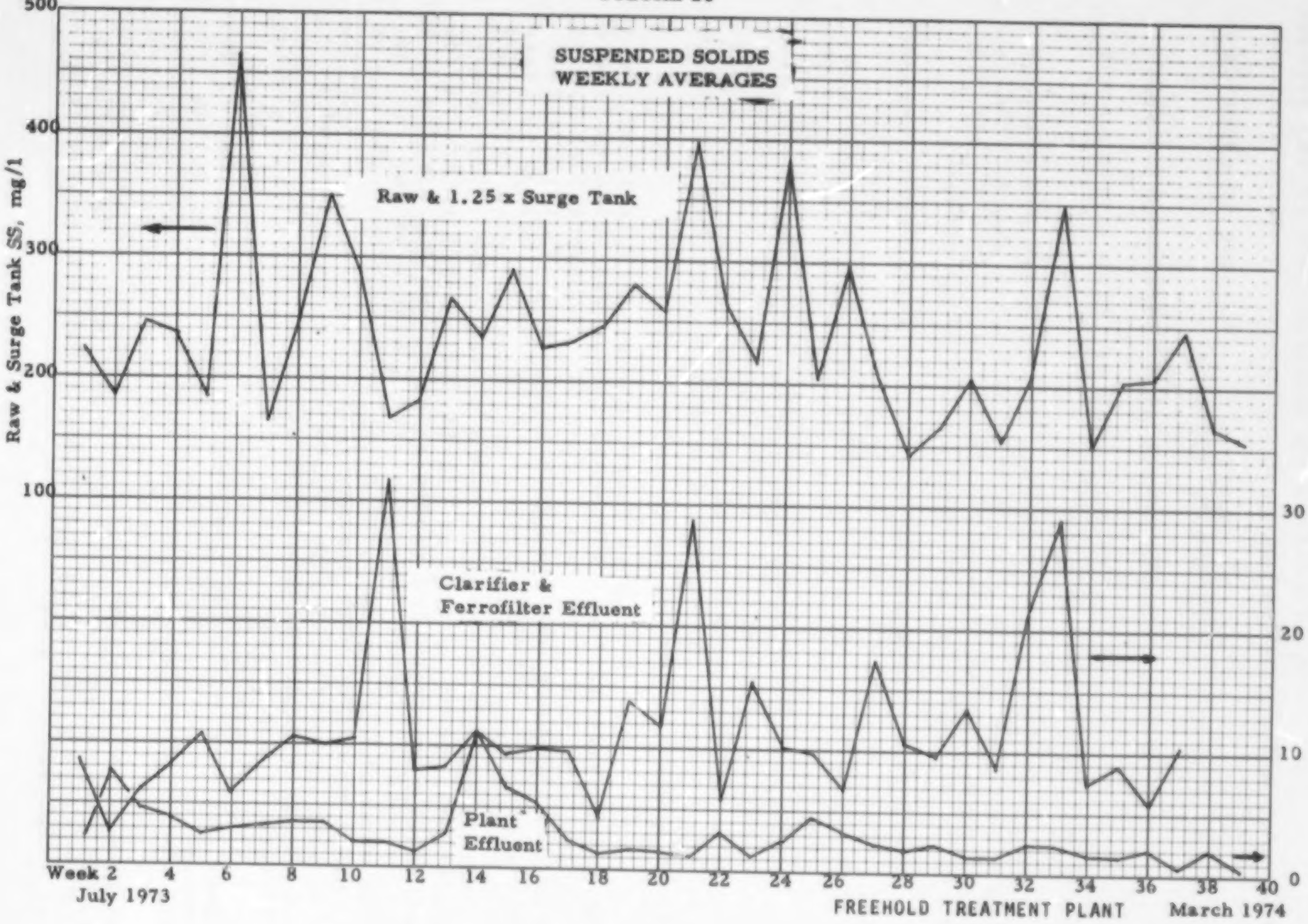

SUSPENDED SOLIDS
WEEKLY AVERAGES
Raw & 1.25 x Surge Tank
Clarifier &
Ferrofilter Effluent
Plant
Effluent
Raw & Surge Tank SS, mg/l
500
400
300
200
100
30
20
10
0
Week 2
4
6
8
10
12
14
16
18
20
22
24
26
28
30
32
34
36
38
40
July 1973
FREEHOLD TREATMENT PLANT
March 1974

FIGURE 27

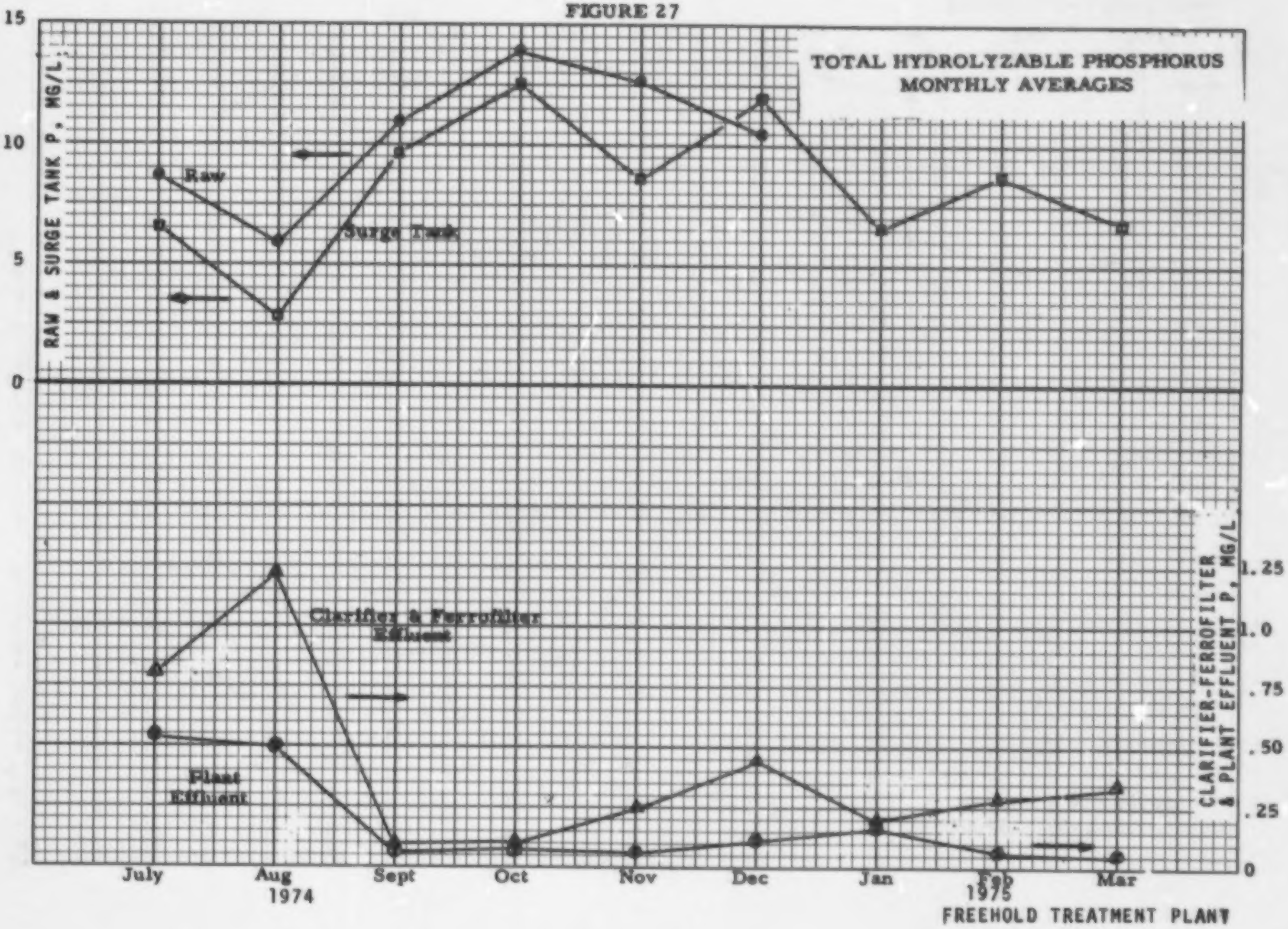

TOTAL HYDROLYZABLE PHOSPHORUS
MONTHLY AVERAGES
RAW & SURGE TANK P, MG/L
15
10
5
0
Raw
Surge Tank
Clarifier & Ferrofilter Effluent
Plant Effluent
CLARIFIER-FERROFILTER & PLANT EFFLUENT P, MG/L
1.25
1.0
.75
.50
.25
0
July
Aug
1974
Sept
Oct
Nov
Dec
Jan
Feb
1975
Mar
FREEHOLD TREATMENT PLANT

FIGURE 28

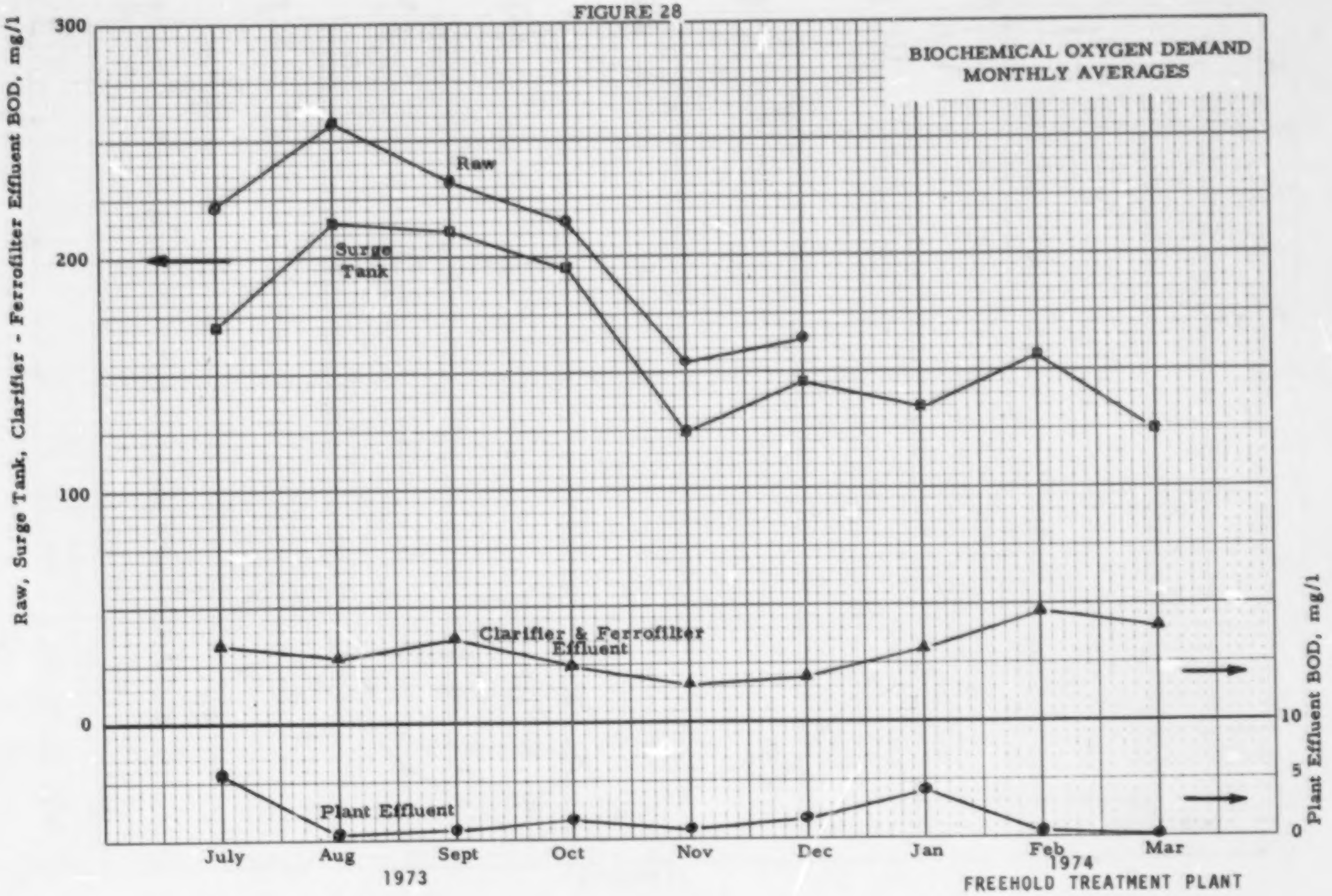

BIOCHEMICAL OXYGEN DEMAND
MONTHLY AVERAGES
Raw, Surge Tank, Clarifier - Ferrofilter Effluent BOD, mg/l
300
200
100
0
Raw
Surge Tank
Clarifier & Ferrofilter Effluent
Plant Effluent
Plant Effluent BOD, mg/l
10
5
0
July
Aug
Sept
Oct
Nov
Dec
Jan
Feb
Mar
1973
1974
FREEHOLD TREATMENT PLANT

FIGURE 29

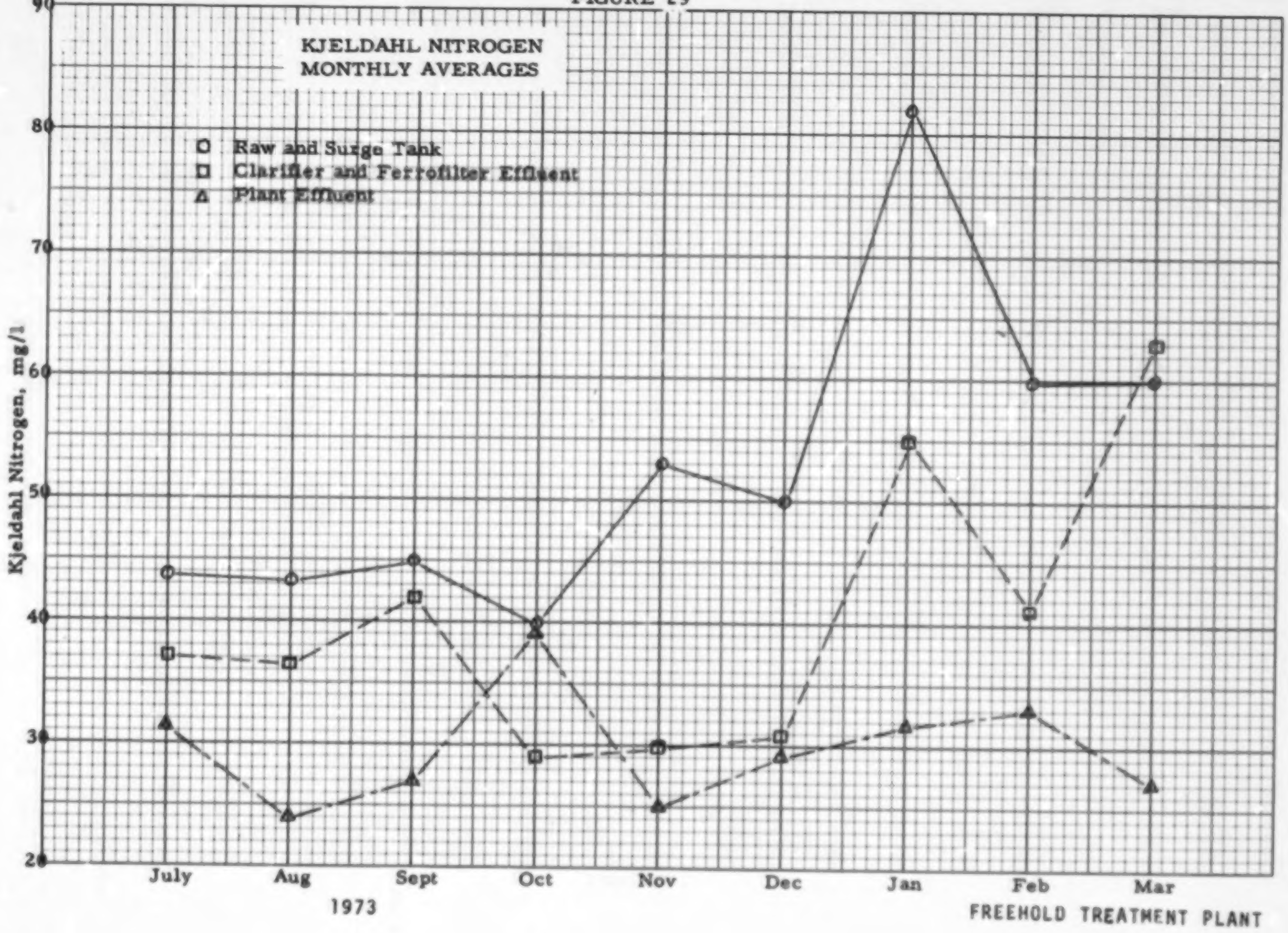

KJELDAHL NITROGEN
MONTHLY AVERAGES
○ Raw and Surge Tank
□ Clarifier and Ferrofilter Effluent
△ Plant Effluent
Kjeldahl Nitrogen, mg/l
90
80
70
60
50
40
30
20
July
Aug
Sept
Oct
Nov
Dec
Jan
Feb
Mar
1973
FREEHOLD TREATMENT PLANT

FIGURE 30

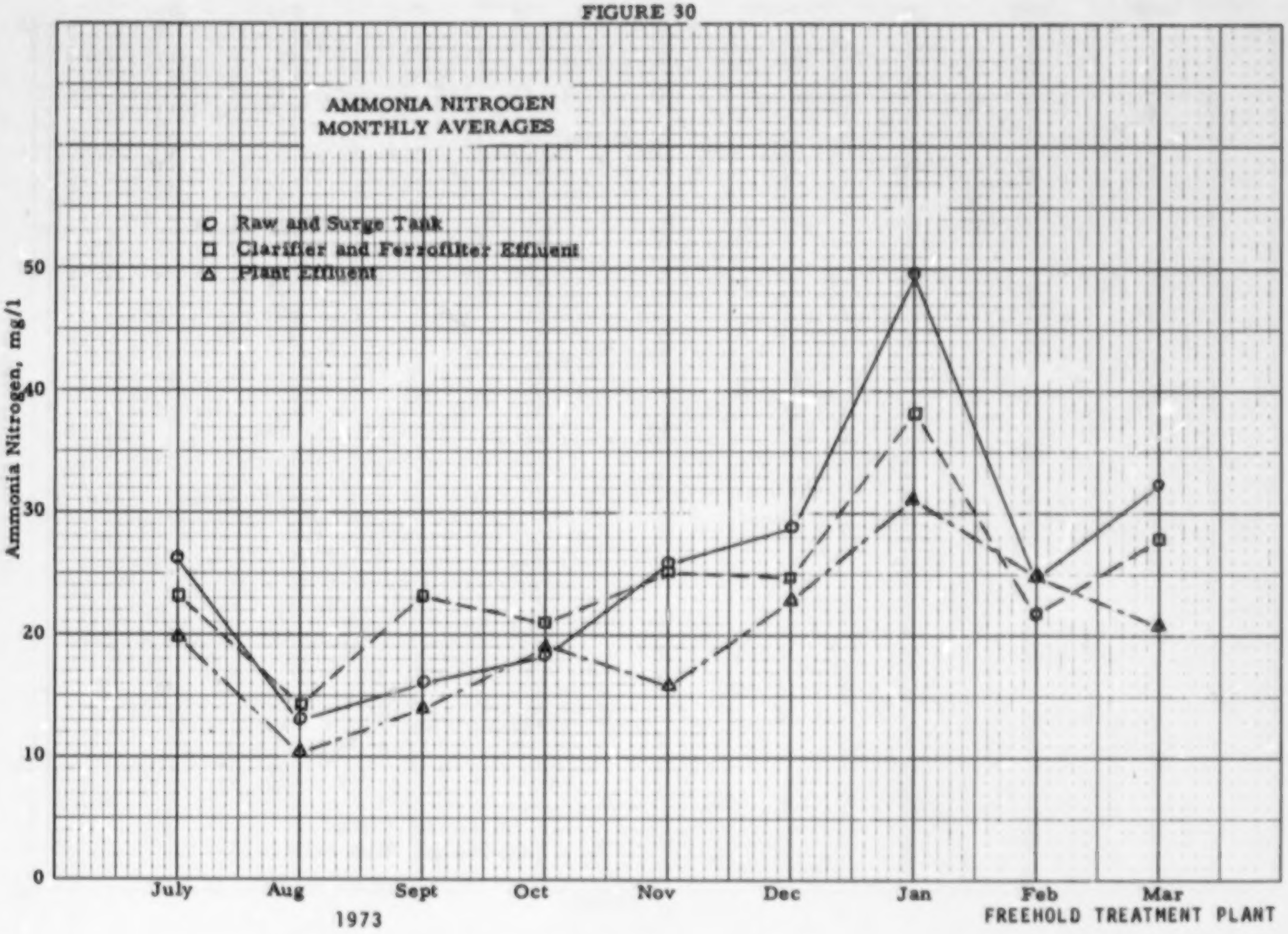

AMMONIA NITROGEN
MONTHLY AVERAGES
○ Raw and Surge Tank
□ Clarifier and Ferrofilter Effluent
△ Plant Effluent
Ammonia Nitrogen, mg/l
0
10
20
30
40
50
July
Aug
Sept
Oct
Nov
Dec
Jan
Feb
Mar
1973
FREEHOLD TREATMENT PLANT

FIGURE 31

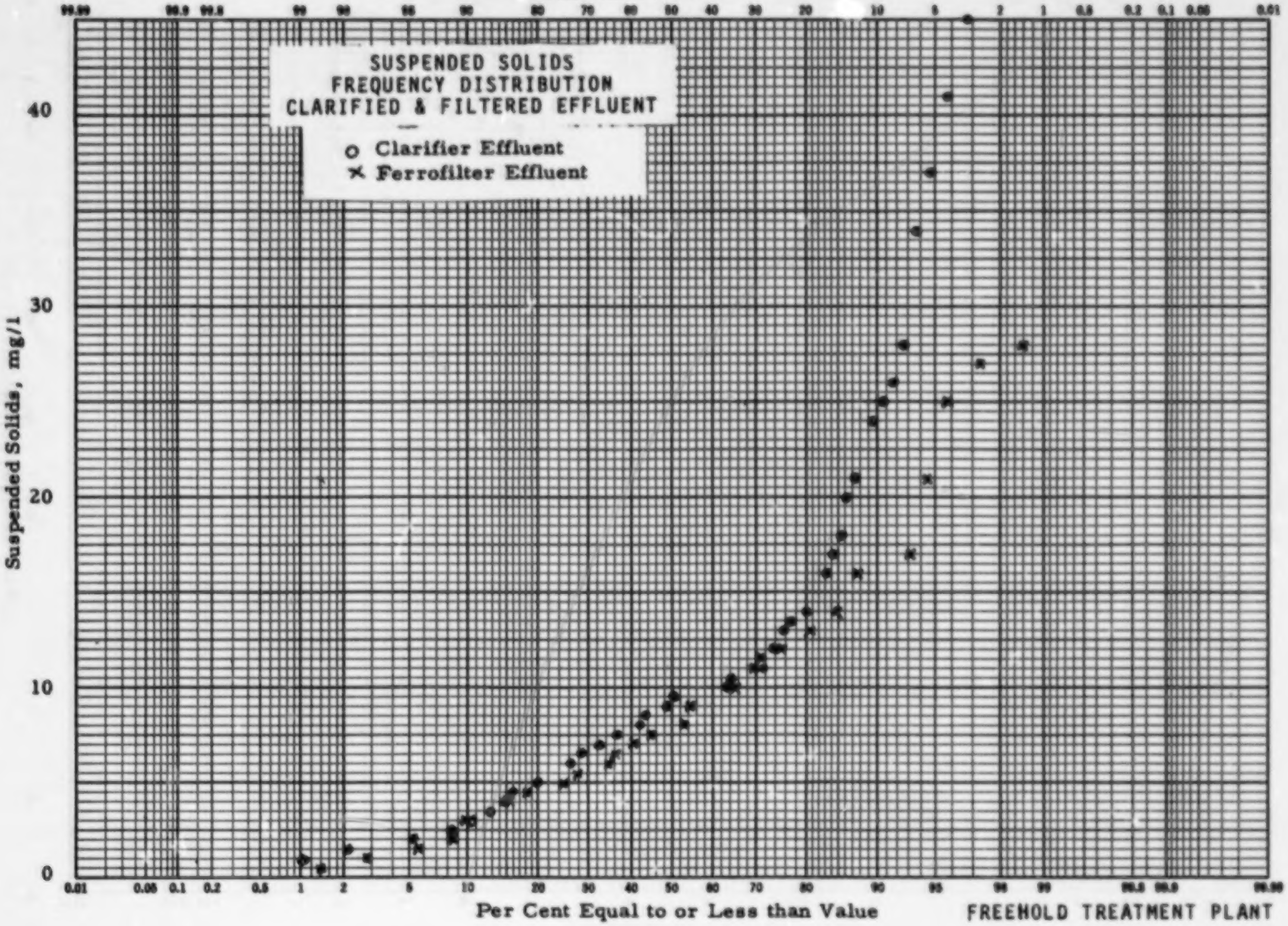
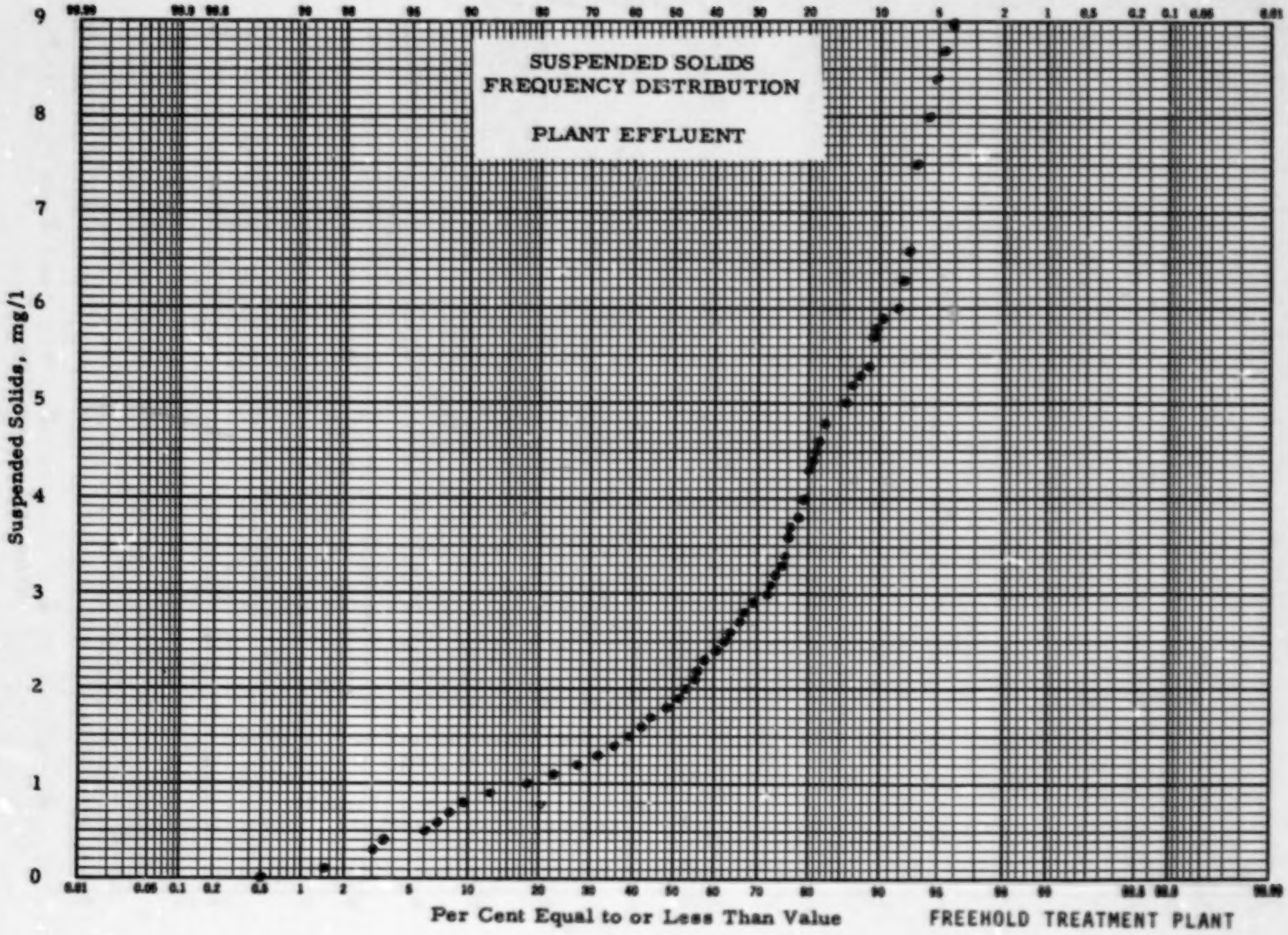

SUSPENDED SOLIDS
FREQUENCY DISTRIBUTION
CLARIFIED & FILTERED EFFLUENT
o Clarifier Effluent
× Ferrofilter Effluent
Suspended Solids, mg/l
0
10
20
30
40
0.01
0.05
0.1
0.2
0.5
1
2
5
10
20
30
40
50
60
70
80
90
95
98
99
99.8
99.9
99.99
Per Cent Equal to or Less than Value
FREEHOLD TREATMENT PLANT

FIGURE 32


SUSPENDED SOLIDS
FREQUENCY DISTRIBUTION
PLANT EFFLUENT
Suspended Solids, mg/1
0
1
2
3
4
5
6
7
8
9
0.01
0.05
0.1
0.2
0.5
1
2
5
10
20
30
40
50
60
70
80
90
95
98
99
99.8
99.9
99.99
Per Cent Equal to or Less Than Value
FREEHOLD TREATMENT PLANT

Ferrofilter effluent BOD (Figure 33) follows a normal frequency distribution, with a median of 30 to 35 mg/l and 95% of the values less than 65 mg/l.

The distribution for plant effluent BOD (Figure 34) is highly skewed, with 62% of the values reported as zero, 90% less than 4, and 99% at 8 mg/l or less. BOD determination in this range are subject to substantial error. The zero values were reported when the seed-corrected oxygen depletion was zero or a negative number. The minimum detectability possible with the method and dilutions used was 0.3 mg/l.

(3) Chemical Consumptions

Chemical consumptions in mg/l (Table 14) were calculated on a monthly basis from raw material receipts, inventory changes, and total effluent flow from the plant. Consumptions during the period of steady state operation were substantially higher than anticipated from previous pilot work and from raw sewage analyses and bench scale tests at the plant. Several factors contributed to these high values:

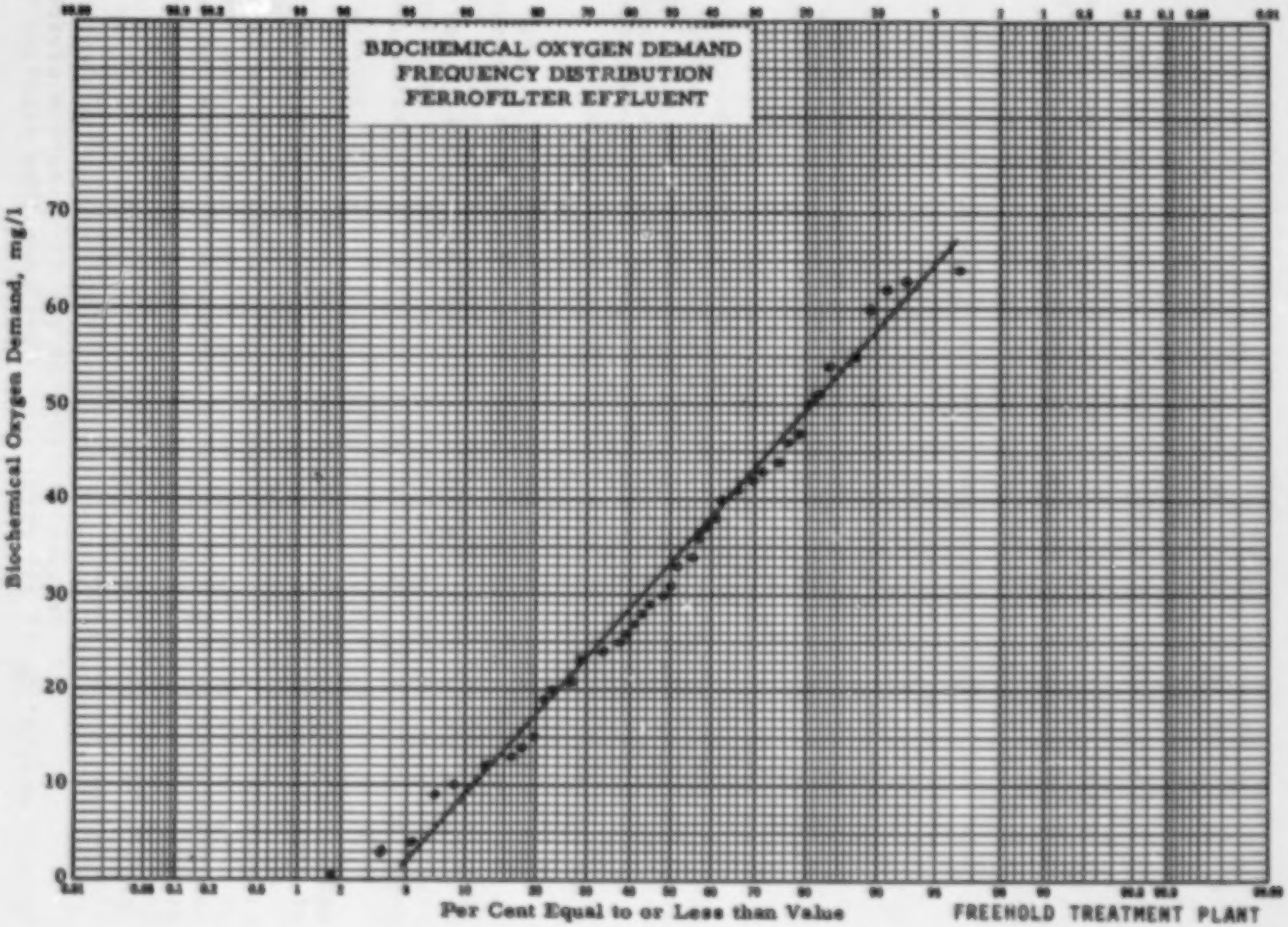
(a) The turbidimetric system designed to control ferric chloride addition was ineffective because of excessive delay in signal response at the flow rates encountered and also because of background interferences within the flocculation and clarification system. This method of control was abandoned early in the operation and replaced with manual adjustment by the operator.

(b) As reported previously, the static mixers in the chemical addition system were removed because of frequent fouling. Although chemical mixing and flocculation appeared to be satisfactory without the mixers, it seems highly likely that their absence contributed to increased requirements of both ferric chloride (with resulting higher NaOH usage) and polyelectrolyte.

(c) Manual control of ferric chloride addition undoubtedly accounted for a large portion of the excess usage of both ferric chloride and sodium hydroxide. Operator surveillance was generally limited to normal daylight hours, the period of maximum ferric chloride demand. A satisfactory rate of ferric chloride addition during these hours was higher than needed for the later evening and early morning hours when phosphate and suspended solids levels were at a minimum. The operator's reluctance to risk an underdose was normal, particularly in view of the emphasis which was placed on obtaining good operation of the liquid handling process.

(d) Frequent failure of the lift station caused excess consumption of all chemicals. This condition resulted in

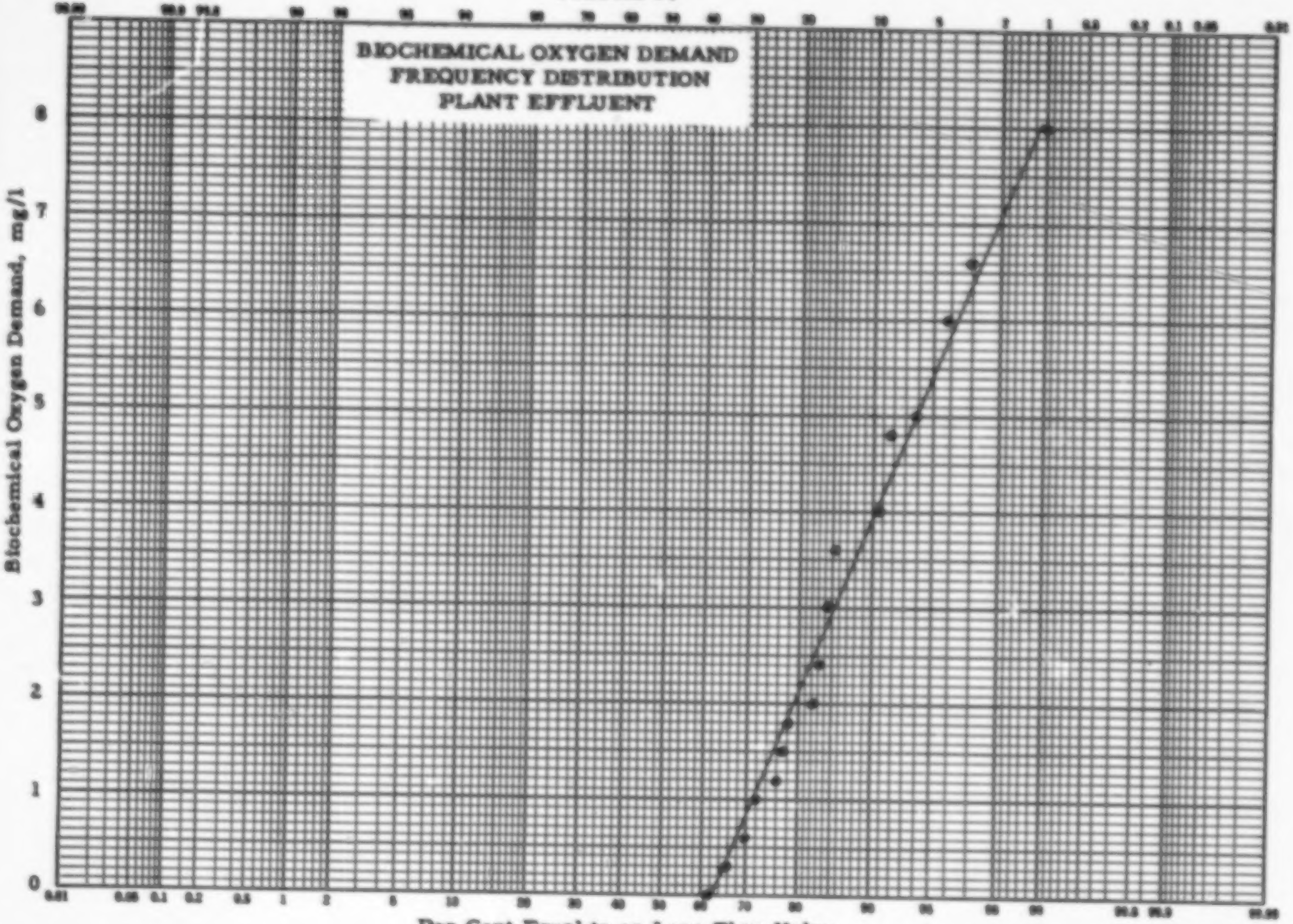
FIGURE 33



BIOCHEMICAL OXYGEN DEMAND
FREQUENCY DISTRIBUTION
FERROFILTER EFFLUENT

FREEHOLD TREATMENT PLANT

FIGURE 34



BIOCHEMICAL OXYGEN DEMAND
FREQUENCY DISTRIBUTION
PLANT EFFLUENT

FREEHOLD TREATMENT PLANT

TABLE 14

CHEMICAL CONSUMPTION

mg/l

Month	$FeCl_3$	NaOH	Hercofloc	Fe_3O_4	Chlorine
July	325	201	2.6	4.5	24.1
Aug.	180	178	2.9	4.0	28.4
Sept.	279	102	3.6	6.1	18.6
Oct.	269	174	2.9	3.6	22.3
Nov.	280	178	1.5	5.1	14.2
Dec.	223	168	1.9	8.3	20.6
Jan.	252	134	2.7	0.4	31.8
Feb.	276	81	2.0	-	34.4
Mar.	232	166	3.1	-	43.4
Avg.	267	154	2.6	5.3 (a)	26.4

(a) Six Months - July through December

a low surge tank level, with automatic recycle of clarifier effluent to the surge tank in order to maintain an operating liquid level. Lift station failure during the day could be corrected by manually operating the lift pumps, but malfunction at other times caused repeat addition of chemicals to the recycling sewage.

(e) On the average the liquid handling system produced a high quality effluent with relatively little difficulty. The need to optimize chemical consumptions, however, was overshadowed by preoccupation with the solids disposal operation which posed a number of difficult problems, as will be discussed later.

(f) High chlorine consumption was due in part to its use to control blockage in the carbon column. Also, the degree of chlorination applied to the plant effluent was probably in excess of that required to meet state standards. As with the other chemical additions the philosophy was generally "better to be safe with a bit too much than not enough."

In summary, chemical consumptions could undoubtedly have been reduced with more work and greater surveillance. A detailed analysis would be required to determine the optimum economic balance among costs of chemicals, plant size, automation and instrumentation to optimize consumptions, and operating and maintenance labor.

Activated carbon usage could not be properly evaluated during this study. Although no thermal carbon regeneration was performed it was necessary, during the course of the evaluation, to transfer carbon from the column for washing, followed by column backwash and reloading. Some carbon was lost physically during these operations (described elsewhere in the report), and was replaced by virgin carbon. Several samples of "spent" carbon taken during these transfers were analyzed. None showed evidence of impending exhaustion. Most exhibited capacities (to adsorb sewage TOC and BOD) of 60 to 80% relative to virgin carbon.

(4) Operation & Maintenance

During the "steady state" mode, one operator was normally assigned to the plant for eight hours a day, Monday through Friday. In addition to making the required operating adjustments, this operator handled the routine maintenance, conducted the analyses necessary to monitor the performance of the plant, and did the housekeeping to keep the building and grounds neat and clean. Weekend coverage was usually limited to the time required to check the operation, collect samples, and run the minimum analyses.

The vacuum lift station continued to fail at irregular intervals due to objects hanging in the check valves and preventing them from closing when the pumps stopped. The liquid would then drain from the lines and the pump prime would be lost. This problem was not solved during the report period; however, at a later date, submersible pumps were installed in place of the vacuum lift pumps.

Periodically during unattended operation when the lift station failed or when the sewage flow was appreciably less than expected, the surge tank would go empty. When this happened the process sewage flow would drop to zero and ferric chloride would back up into the steel process line resulting in severe corrosion. When flow resumed, it was often irregular for a time and the clarifier operation would be upset. A relay was installed on the surge tank level recorded which, when the level dropped to 13% would activate the existing automatic valve in the clarifier effluent line to recycle the flow to the surge tank, and when the surge tank level increased to 22% would reset the recycle valve for normal operation. The system then operated smoothly in the event of lift station failure or low influent sewage flow, however, chemical consumption was increased.

As explained elsewhere, manual control of ferric chloride addition accounted for a large portion of the excess use of chemicals. It was observed that good flocculation was consistantly obtained when sufficient ferric chloride was added to depress the pH to about 5.9 to 6.3. A pH detector-controller was installed and connected so as to regulate ferric chloride addition to maintain the set pH of about 6.1. Operation was satisfactory, however, the instrument failed after ten days operation due to a defective seal. It was returned to the manufacturer and not reinstalled during the report period.

The clarifier rake drive bearings failed on July 8, and were replaced on July 10. An inspection revealed that one of the original bearings supplied in the gear box was defective. Clarifier effluent turbidity was normal at 3 to 6 JTU during the two-day period without rake operation, however, the sludge pumped to the hold tank was very thin, less than 2% total solids.

The Ferrofilter was a major power consumer yet its efficiency in removing suspended solids was low. In November the unit was opened and the grids were removed, cleaned, and replaced. It was estimated that they were 70 to 80% plugged. There was little improvement so on January 17, use of the unit was discontinued.

The four woven wire effluent screens at the top of the adsorber required cleaning every two weeks. During October, they were replaced with wedge-wire well point screens. These worked very well. The cleaning frequency was reduced to once per month. Screen boxes were fabricated and installed on both the adsorber and quench tank drain lines to catch any carbon that might be lost when backwashing or draining these units.

An occasional problem with poor flocculation in the clarifier and high Hercofloc consumption was believed to be caused by non-uniform Hercofloc solutions. The mixing of polymer and water by the original small agitators was inadequate. Double bladed agitator shafts driven by 1/4 H.P. motors were provided in January, 1974. The Hercofloc solution and flocculation seemed to be improved.

Plant operation during the "steady state" period was generally good and effluent quality was consistantly high. Only two major deviations from the design concept were made. First, the Ferrofilter was removed from operation as described earlier due to poor efficiency and high power consumption. Second, the carbon was not thermally regenerated because by the end of the report period, it still had a capacity to adsorb BOD of at least 60% relative to virgin carbon.

b. Intensive Evaluation I

Sampling for intensive evaluation was conducted under two different plans. The first, which was shown in Table 9 (Normal Evaluation Section), specified three sets of samples for each of two days per week for eight weeks. The three samples consisted of composites taken during the following periods:

Period	Hours
1	0800 - 1400
2	1400 - 2000
3	2000 - 0800

The following six locations were selected with the primary objective of determining parameter variations during the three sampling periods:

Raw - Lift Station Discharge to Hydrasieve Box

ST - Discharge from Surge Tank

CE - Clarifier Effluent

FFE - FerroFilter Effluent

CCE - Carbon Column Effluent

PE - Plant Effluent

Generally, eleven separate parameters were analyzed with emphasis stressed primarily on the Raw, Surge Tank, and Plant Effluent locations.

The above evaluation was terminated at the end of the fifth week instead of the scheduled eighth week. It was felt at that point that the trend of the data gathered did not and would not show the kind of changes which were expected to occur during the various times of a typical diurnal cycle. Since most of the other data gathered at this plant showed that there are indeed variations in the measured parameters during the course of the day, it can only be postulated that the time periods chosen were most likely not the best and/or that the number of periods were too few to show any variation.

At the end of the fifth week, approximately 1400 data points had been determined. The data are summarized in Table 15. The value given for each time period is an arithmetic mean of approximately ten determinations. A few values were missing and several were discarded as being extremely outside the normal population. The unused and missing numbers comprised slightly over 1% of the total number of determinations.

Observation

As previously stated, the data accumulated did not follow the trend expected and must therefore be carefully scrutinized in order to extract any meaningful conclusions. It was expected that a significant drop in the absolute values of the parameters would be observed during the second and third periods; similar to that displayed by a characteristic diurnal flow pattern.

Generally, the pollutant parameters measured in the Raw Sewage and Surge Tank did show a slight drop in the parameters for the third period with the exception of alkalinity, TKN, and NH_3-N which were appreciably higher than in the other two periods. This sequence, however, does not extend to all the locations sampled. For example, all three periods of the plant effluent showed a random trend with the exception of the TOC, Alkalinity, TKN, and NH_3-N which displayed a significant downward trend.

Another observation, for which there is no apparent explanation, is the large values of total phosphorus in the raw sewage and surge tank. For the most part, the values are three times the expected normal and the tendency would be to assume that the concentrations were expressed as PO_4 and not P. However, all laboratory reports indicate that this is not the case. The soluble phosphorus levels obtained during all three periods do fall within the expected values.

TABLE 15

Intensive Evaluation I

Average Values, mg/l

Period	Raw	ST	CE	FEE	PE
Total Phosphorus					
1	29.1	26.4	1.6	2.4	0.70
2	26.5	23.7	1.6	3.2	0.74
3	26.3	18.8	1.7	2.1	0.66
Average	27.3	23.1	1.7	2.5	0.70
Soluble Phosphorus					
1		8.7		0.83	0.43
2		9.6		0.92	0.39
3		6.4		0.84	0.48
Average		8.4		0.86	0.43
Alkalinity					
1	218	237		179	147
2	202	225		168	143
3	269	246		164	130
Average	230	236		170	140
Coliform (MPN - Geometric Mean)					
1	7.0×10^{5}	5.1×10^{5}	2.7×10^{4}	2.8×10^{4}	2.4
2	7.7×10^{5}	1.2×10^{6}	3.1×10^{4}	2.5×10^{4}	<1
3	6.5×10^{5}	7.2×10^{5}	1.6×10^{4}	1.5×10^{4}	3.0
Average	7.1×10^{5}	7.7×10^{5}	2.4×10^{4}	2.0×10^{4}	2.0

FREEHOLD TREATMENT PLANT

TABLE 15 (cont.)

SS

Period	Raw	ST	CE	FEE	CCE	PE
1	261	304	21.4	17.4	3.9	3.5
2	267	298	18.9	18.8	4.1	4.7
3	276	207	15.3	21.2	4.6	3.1
Average	268	272	18.4	19.2	4.2	3.7
TOC						
1	132	117		26.0	8.5	8.8
2	123	115		25.2	8.5	8.2
3	116	99		23.7	8.9	7.6
Average	124	114		25.0	8.6	8.2
SOC						
1		32.8		19.8	6.3	7.1
2		31.9		18.4	6.4	6.8
3		29.4		18.2	6.0	7.4
Average		31.4		18.8	6.2	7.1
BOD						
1	198	185				5.7
2	233	172				5.7
3	200	146				5.9
Average	210	169				5.8

FREEHOLD TREATMENT PLANT

TABLE 15 (cont.)

TKN

Period	Raw	ST	FFE	PE
1	38.7	42.5	33.7	28.3
2	34.1	37.2	28.9	25.5
3	58.2	49.0	28.9	21.3
Average	43.2	42.7	30.5	25.0

NO_2+NO_3-N

Period	Raw	ST	FFE	PE
1	0.40	0.31	0.13	2.05
2	0.23	0.35	0.14	1.93
3	0.23	0.20	0.11	2.55
Average	0.29	0.29	0.13	2.18

NH_3-N

Period	Raw	ST	FFE	PE
1	23.9	31.8	28.1	29.3
2	23.5	27.0	26.9	24.8
3	47.9	37.6	26.7	21.0
Average	31.8	32.1	26.3	25.0

Six sets of sludge samples were taken during this period. The results are covered in the Solids Disposal section of the Report.

FREEHOLD TREATMENT PLANT

Of interest is the significant increase in the amount of NO_2 + NO_3-N in the plant effluent as compared to pre-carbon column values. These results would indicate a substantial amount of nitri-nitrafication occuring in the carbon column. Such results are not totally unexpected considering the significant amount of bio-activity which is present in the carbon column, and which is also responsible for extending the observed carbon life.

c. Intensive Evaluation, II

The second intensive evaluation was performed from January 17, 1974, to April 7, 1974. In this study, raw sewage and surge tank samples were taken every two hours (with the exception of 3 a.m.) for ten separate 24-hour days. The evaluation's main objective was to generate a 24-hour profile of the raw sewage and surge tank. Each sample taken was analyzed for eight parameters consisting of BOD, SS, Total P, TOC, TOD, TKN, pH, and NH_3-N.

Approximately 1824 determinations were scheduled during the evaluation period. Of this number, 140 determinations (8% of total) were not performed because of missed samples, and 17 (1%) were discarded as being extremely incongruous with the normal population.

Observations

A summary of the data obtained is presented on Table 16 and graphically represented on Figures 35 thru 42.

In general, the data indicate concentration profiles similar to the characteristic diurnal flow pattern as shown in Figure 21 (Normal Evaluation section). The data demonstrate quite vividly the importance of the surge tank and its function in accepting the highly variable concentrations in the raw influent and dampening them into more easily handled fluctuations. It is also interesting to note that the dampening ability of the surge tank is much more pronounced during periods of high flow and high concentration as opposed to lower flow and lower concentration of the raw influent. One of the reasons for such a difference is that lower flows are experienced during the late evening and early morning hours which are also the hours of minimum surge tank level and, therefore, minimum dampening ability.

In Intensive Evaluation I, it was stated that the three time periods chosen may not have been the best to produce the results expected. As an approximate check of this statement, the data summarized in Table 16 was regrouped as shown in Table 17. The numbers were obtained by averaging the hourly samples of Table 16 during the following periods:

TABLE 16

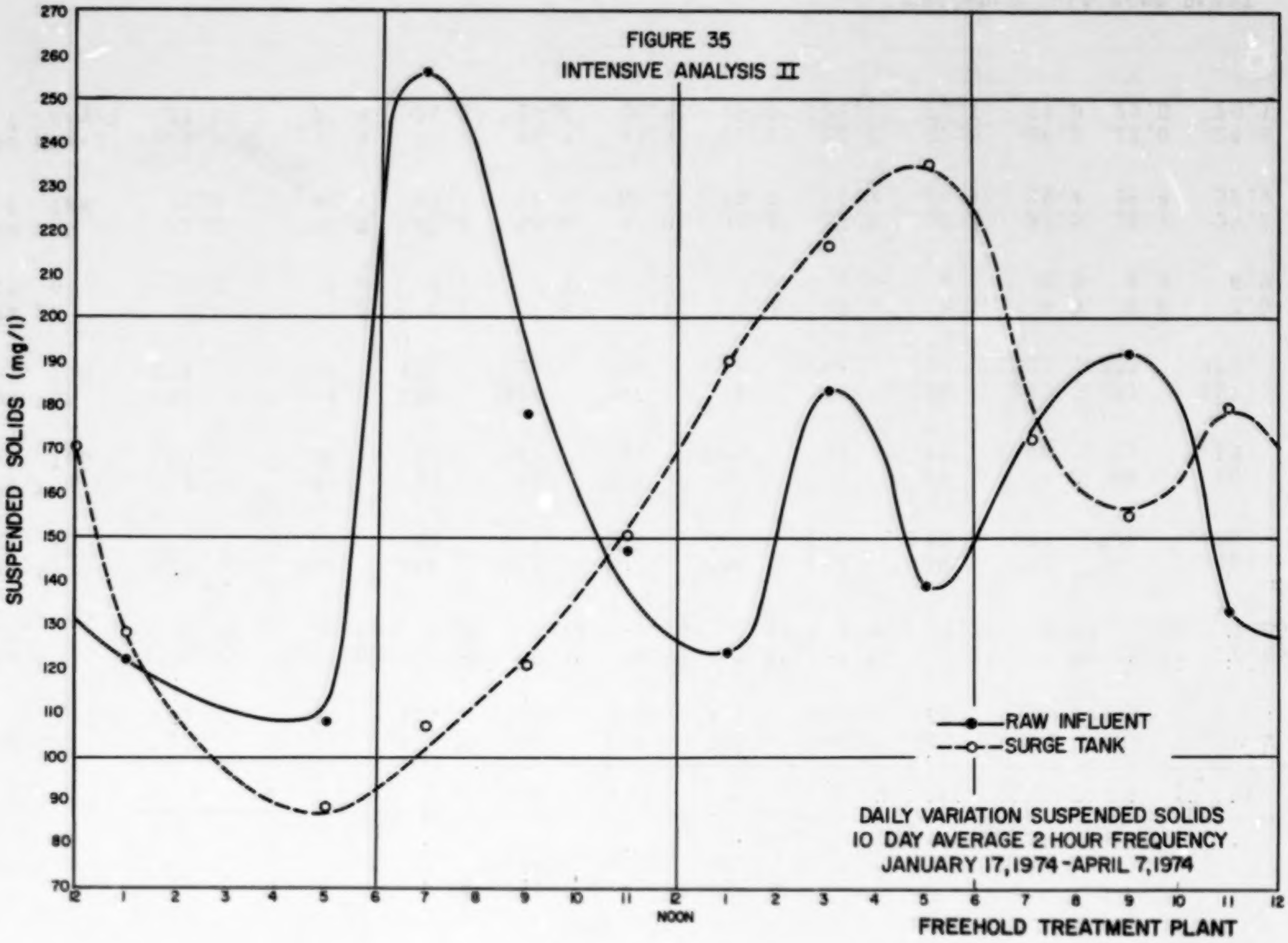
INTENSIVE EVALUATION, II
AVERAGE VALUES, MG/L

	AM						N		PM			
	1	3	5	7	9	11	1	3	5	7	9	11
RAW BOD	137		79	139	169	246	185	146	154	216	244	152
S.T. BOD	163		94	101	127	172	181	159	159	158	179	190
RAW P	5.90		4.60	5.87	8.40	10.51	8.68	8.26	7.38	7.53	6.54	6.80
S.T. P	6.98		6.03	6.04	7.14	8.20	8.51	8.45	7.37	6.97	7.38	7.06
RAW SS	122		107	256	177	187	124	183	139	173	192	133
S.T. SS	128		89	107	121	150	189	216	235	173	156	179
RAW TOC	47		46	54	72	117	83	73	65	77	65	66
S.T. TOC	50		36	40	47	53	60	62	58	56	54	53
RAW TOD	182		229	245	324	397	319	310	225	299	277	251
S.T. TOD	211		186	175	220	248	249	267	235	223	221	221
RAW pH	7.1		7.2	7.7	7.8	7.1	7.0	7.0	7.1	6.9	6.8	7.0
S.T. pH	6.9		6.8	7.1	7.3	7.2	7.0	6.9	6.9	6.9	6.8	6.8
RAW TKN	42.8		48.8	62.8	68.8	47.8	38.2	35.2	36.2	37.5	38.1	35.2
S.T. TKN	38.0		43.1	47.1	56.6	59.2	43.4	35.3	33.9	35.4	38.8	32.9
RAW NH_3-N	25.9		31.9	41.1	45.1	33.4	25.8	26.5	22.6	26.3	22.0	24.6
S.T. NH_3-N	23.5		21.8	25.8	34.6	30.0	29.0	26.0	22.1	24.0	23.0	20.1

FREEHOLD TREATMENT PLANT

FIGURE 35

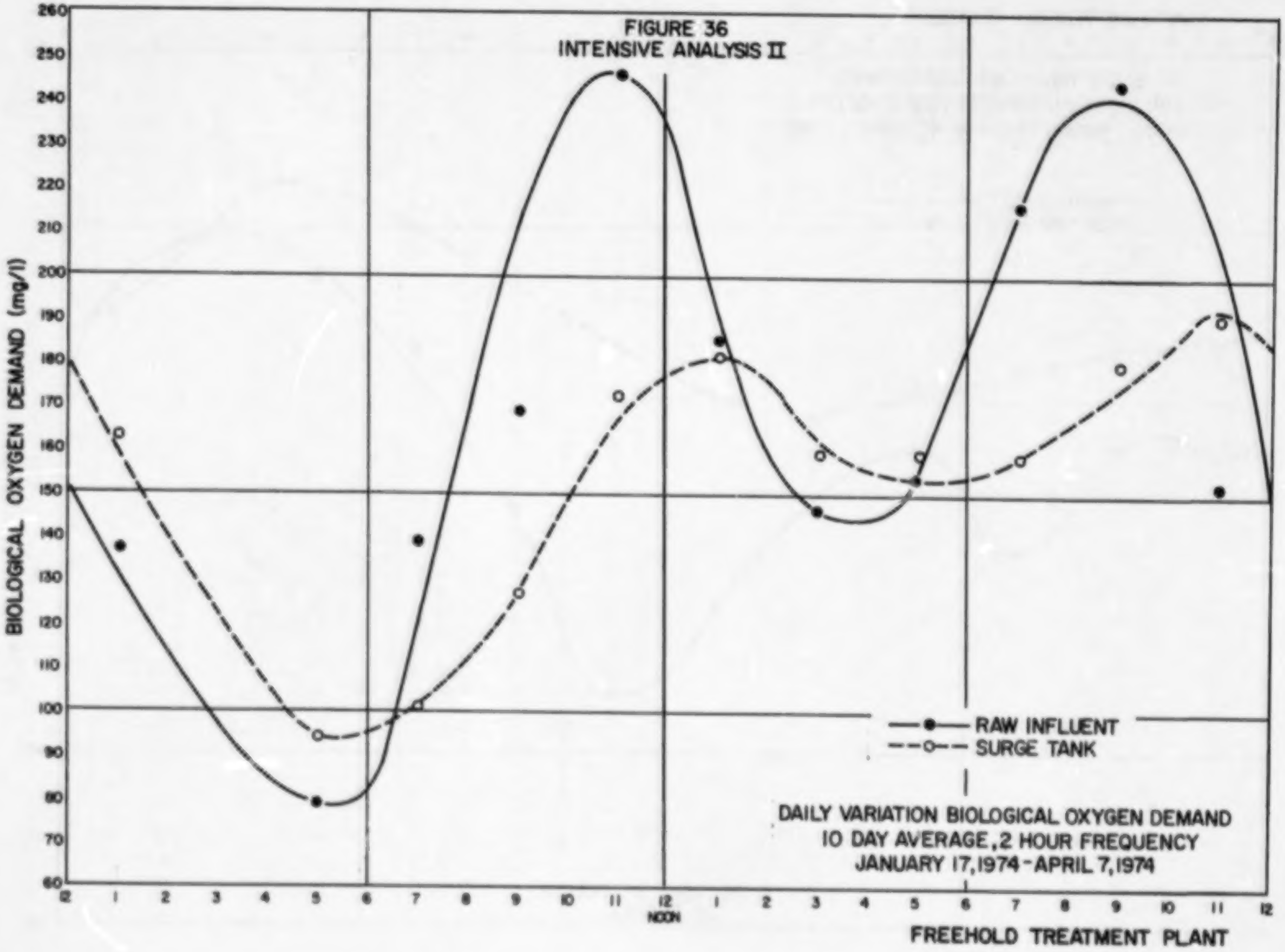
INTENSIVE ANALYSIS II



DAILY VARIATION SUSPENDED SOLIDS
10 DAY AVERAGE 2 HOUR FREQUENCY
JANUARY 17, 1974 - APRIL 7, 1974

FREEHOLD TREATMENT PLANT

FIGURE 36
INTENSIVE ANALYSIS II



DAILY VARIATION BIOLOGICAL OXYGEN DEMAND
10 DAY AVERAGE, 2 HOUR FREQUENCY
JANUARY 17, 1974 - APRIL 7, 1974

FREEHOLD TREATMENT PLANT

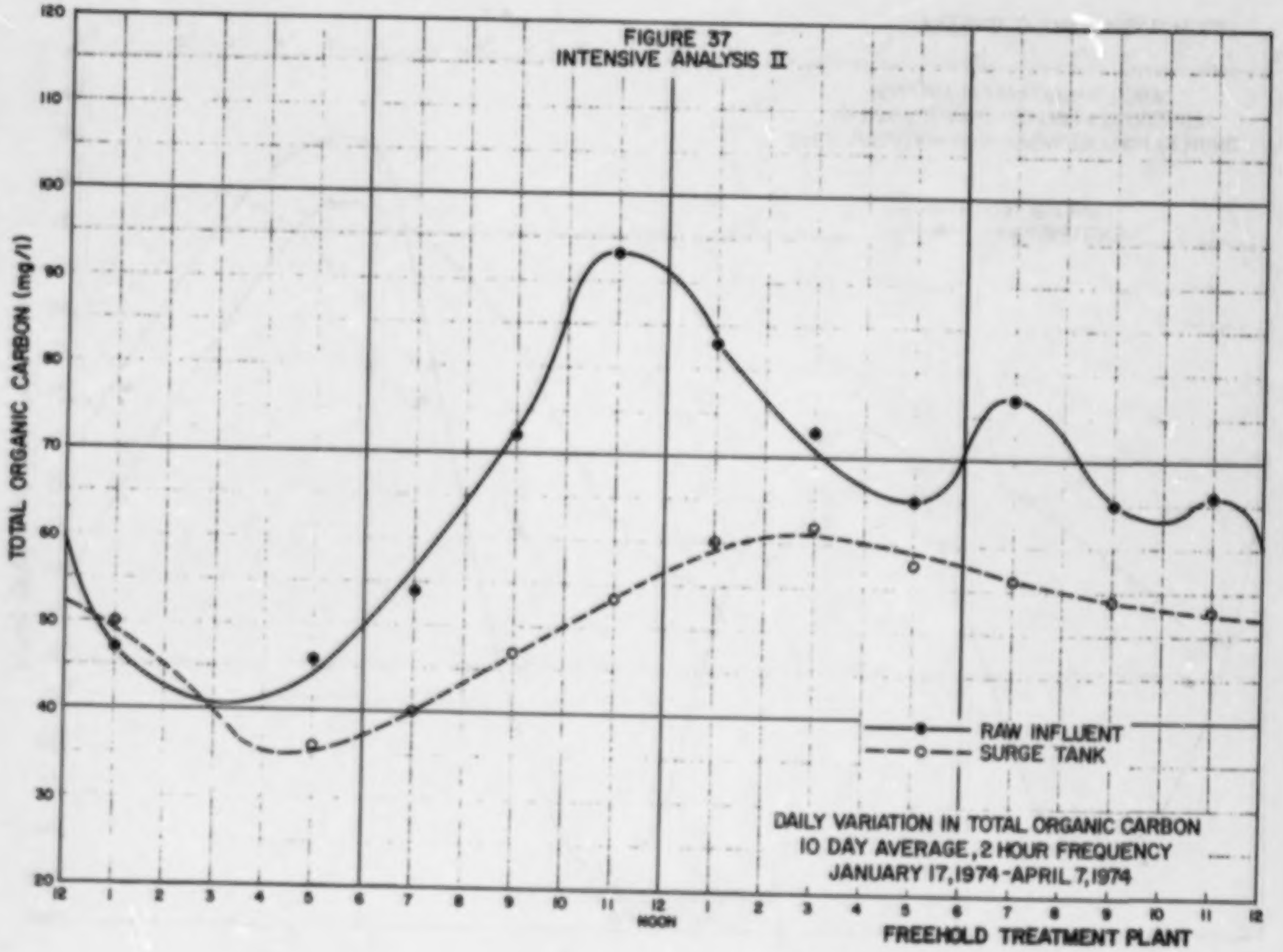
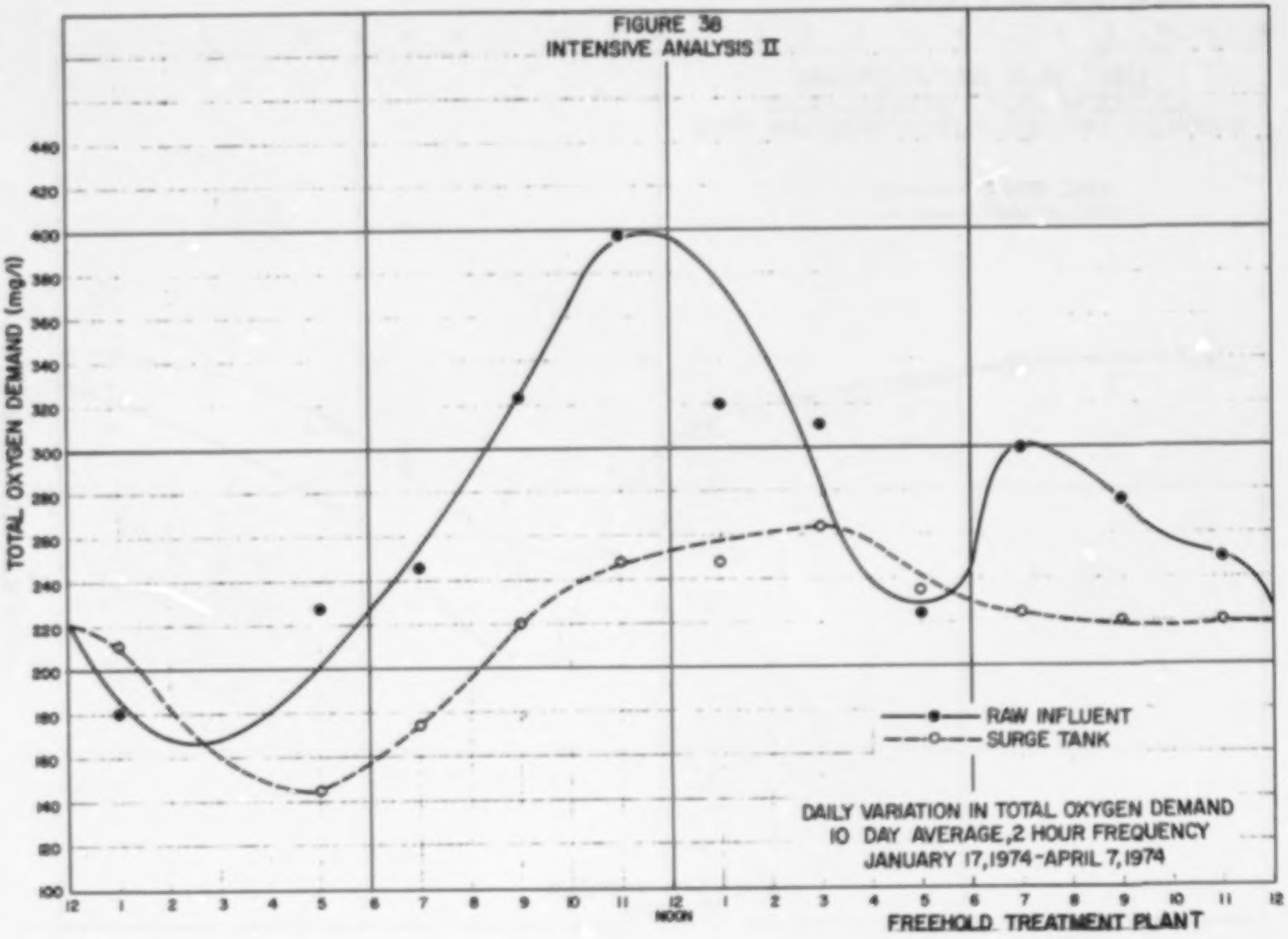


FIGURE 37
INTENSIVE ANALYSIS II

FIGURE 38
INTENSIVE ANALYSIS II



DAILY VARIATION IN TOTAL OXYGEN DEMAND
10 DAY AVERAGE, 2 HOUR FREQUENCY
JANUARY 17, 1974-APRIL 7, 1974

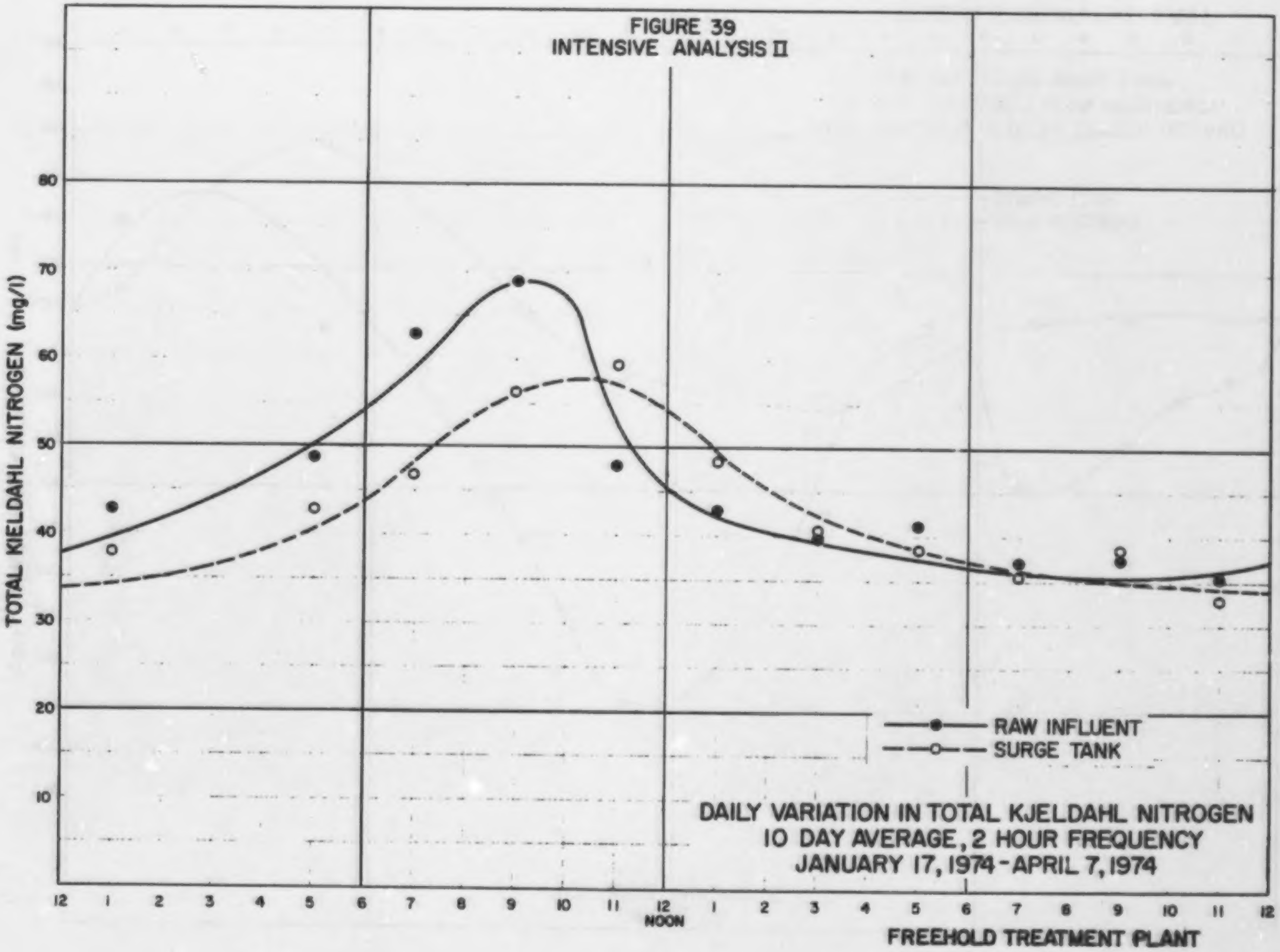
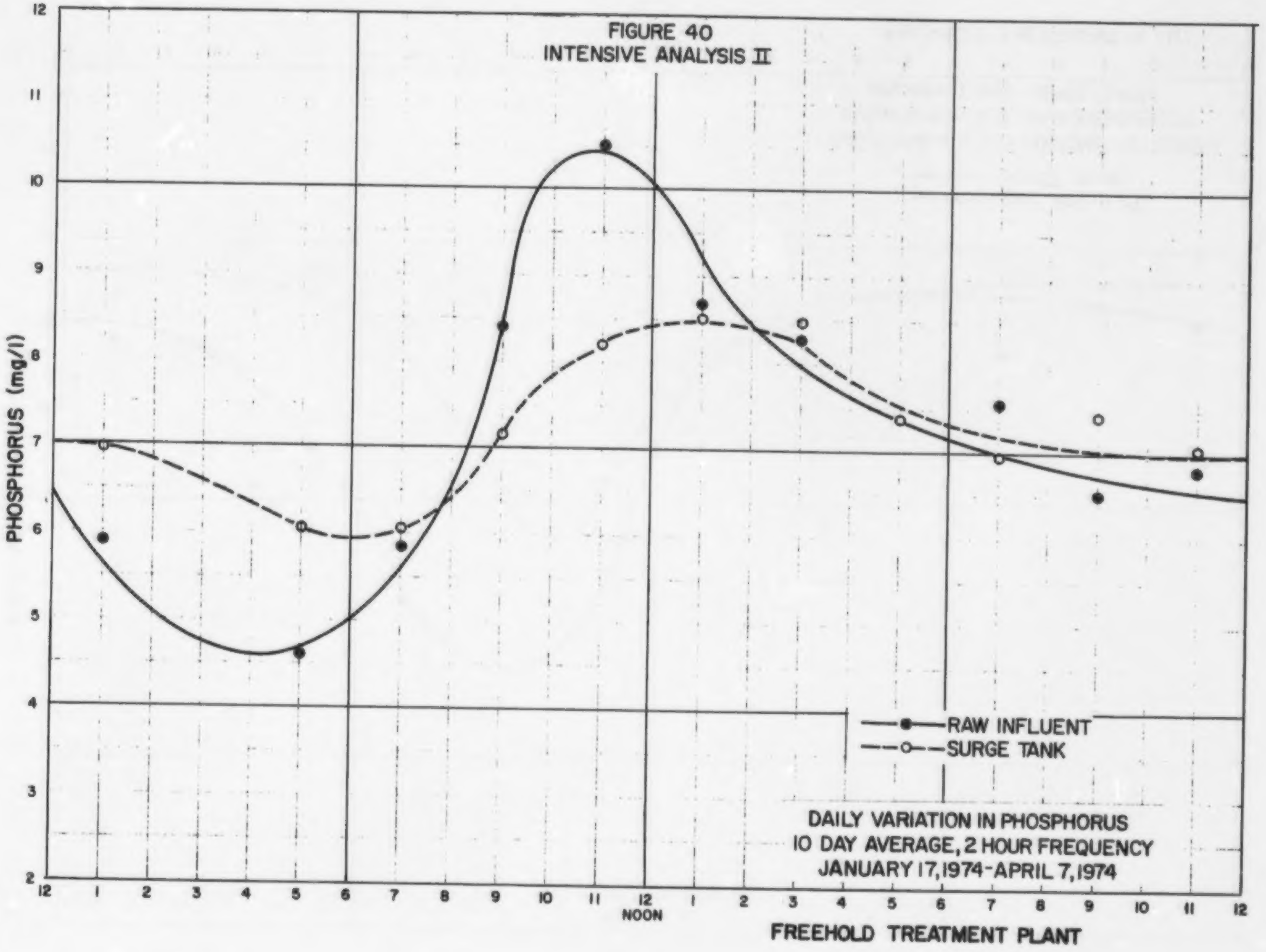


FIGURE 39
INTENSIVE ANALYSIS II

DAILY VARIATION IN TOTAL KJELDAHL NITROGEN
10 DAY AVERAGE, 2 HOUR FREQUENCY
JANUARY 17, 1974 - APRIL 7, 1974

FREEHOLD TREATMENT PLANT

FIGURE 40
INTENSIVE ANALYSIS II



DAILY VARIATION IN PHOSPHORUS
10 DAY AVERAGE, 2 HOUR FREQUENCY
JANUARY 17, 1974-APRIL 7, 1974

FREEHOLD TREATMENT PLANT

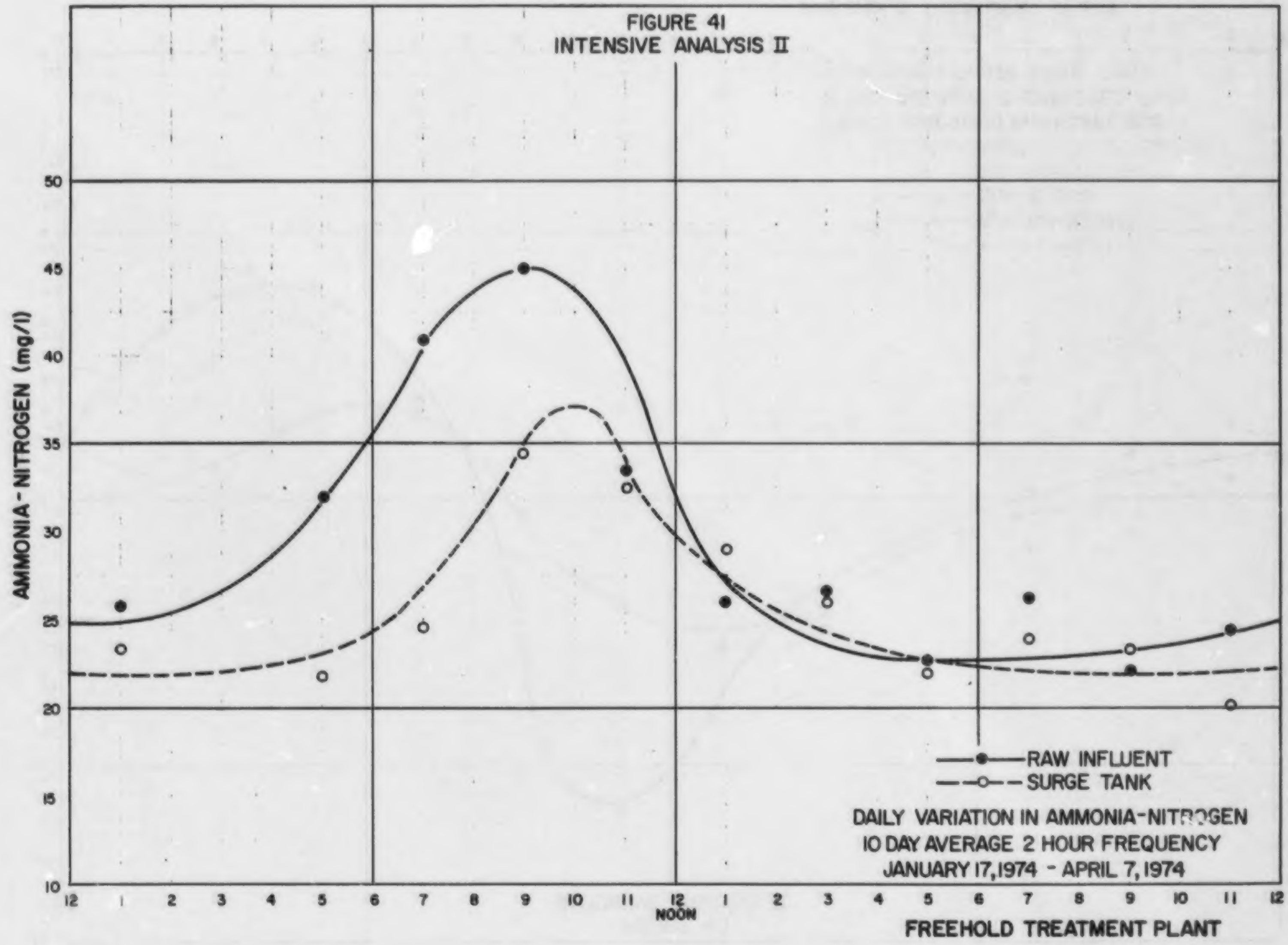


FIGURE 41
INTENSIVE ANALYSIS II

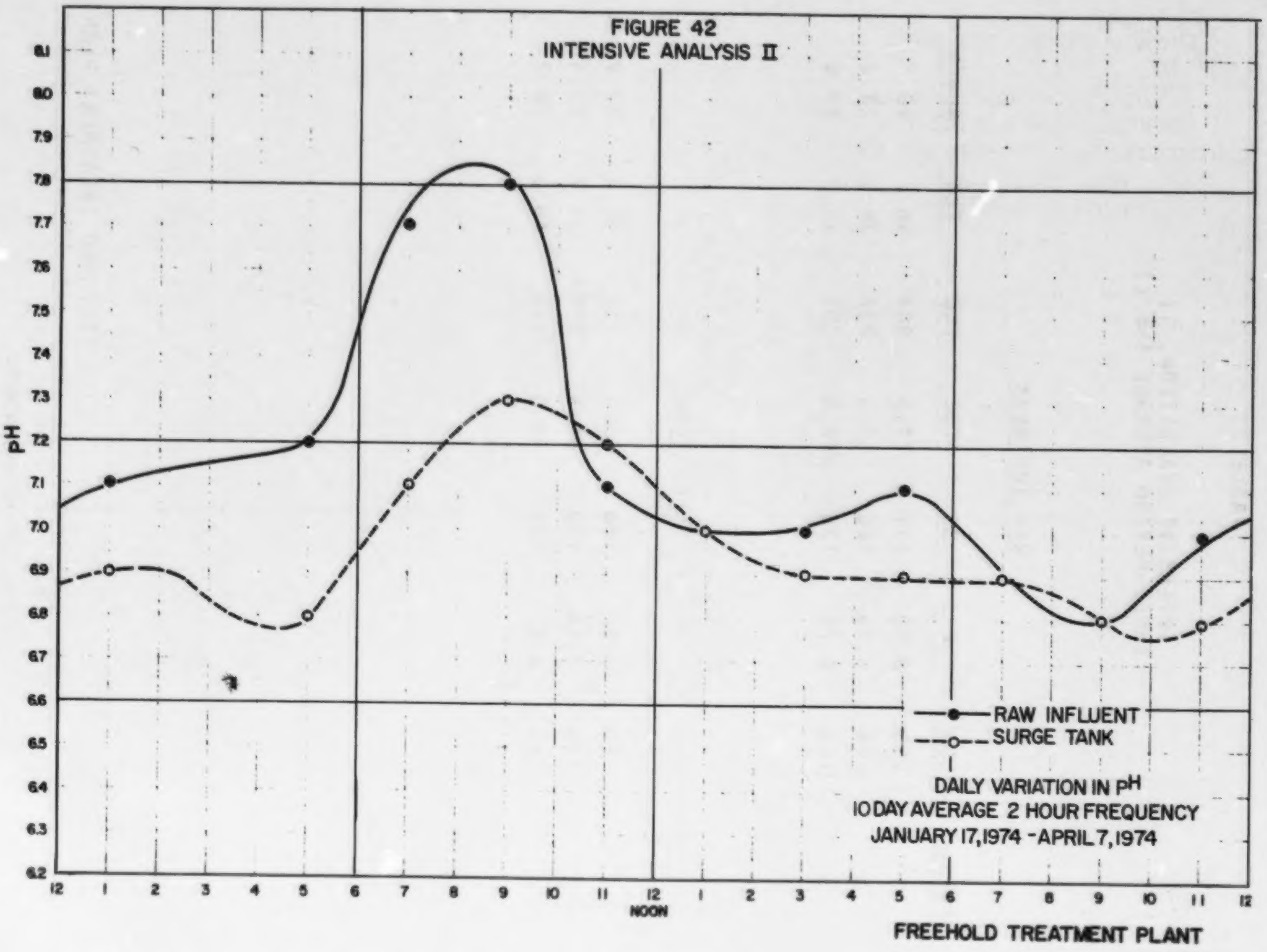


FIGURE 42
INTENSIVE ANALYSIS II

DAILY VARIATION IN pH
10 DAY AVERAGE 2 HOUR FREQUENCY
JANUARY 17, 1974 - APRIL 7, 1974

FREEHOLD TREATMENT PLANT

TABLE 17

INTENSIVE EVALUATION, II
THREE-PERIOD AVERAGE (MG/L)

Raw Influent

Period	BOD	P	SS	TOC	TOD	TKN	NH_3-N
1	255	8.18	171	77.3	304	48.2	40
2	204	6.96	166	69.3	276	36.9	24.3
3	108	5.25	115	46.5	206	45.8	28.9

Surge Tank

Period	BOD	P	SS	TOC	TOD	TKN	NH_3-N
1	150	7.62	170	53.3	232	46.0	27.8
2	176	7.14	169	54.4	222	35.7	22.4
3	129	6.51	109	43.0	176	40.0	22.7

FREEHOLD TREATMENT PLANT

Period	Hours
1	6 a.m. - 6 p.m.
2	6 p.m. - 12 midnight
3	12 midnight - 6 a.m.

This regrouped data indicates a substantial drop in all the parameters during the second and third periods of operation with the exception of TKN and NH_3-N. Both of these parameters, however, did experience a substantial drop during the second period. Evaluation of concentration changes occuring during the day is important in evaluating chemical usage, particularly when chemicals are being added manually. For example, Figure 40 shows significantly less phosphorus in the late evening and early morning periods than during the 6 a.m. to 6 p.m. period; however, the rate of $FeCl_3$ use, which is directly related to the phosphorus content, is normally controlled at an amount which corresponds to that required during the late morning-early afternoon periods. This action, by the operator, of not resetting his chemicals for the late evening-early morning period is understandable since he does not wish to experience a possible upset in the early evening when he has left.

Since the treatment plant's sewage collection system is a relatively tight one, it would be reasonable to propose the following periods of concentration variations:

Five Peaks

-Early morning shower, breakfast, and grooming
-Middle morning cleaning and cloth washing
-Lunch
-Supper
-Retiring

This, of course, did not occur during the intensive evaluation period.

Observation of Figures 35 thru 42 do show, however, that each parameter experiences at least one major peak (with the exception of BOD in which the raw experiences two major peaks) and a few minor peaks.

The major peak appears to begin at approximately 5 a.m. and continues until the middle afternoon. The first minor peaks occur from 5 to 9 p.m., with the last peak beginning at 9 a.m. and lasting only until 11 p.m. at which time all the parameters begin their downward plunge. It seems that of the five peaks proposed, the first three overlap to such an extent that they appear to be one.

It is interesting to note that all parameters did not experience similar peaks during the day. For example, TOD, TOC, SS, pH, and BOD experienced more than one significant peak while TKN, P, and NH_3-N did not. The significance of this is not quite understood at present.

There is also some doubt as to the correctness of the suspended solids curve (Figure 35). The time interval between the raw and surge tank peaks seem rather long. The rapid increase and decrease in the raw should have produced a similar narrow peak in the surge tank curve. The surge tank curve did not display this peak. It also showed very little dampening ability.

B. SOLIDS DISPOSAL

1. Sludge Characteristics

During the first six months of start-up and break-in operation (October 1972-March 1973) the total solids content of the clarifier sludge ranged from about two to four percent with three percent being a typical value. The reason, as described previously was due to erratic operation of the sludge level controller which frequently allowed very thin sludge to be pumped to the Sludge Hold Tank. It was observed that sludge in the hold tank would settle leaving a relatively clear layer of water on top. Five nozzles were welded on the hold tank, equally spaced from the bottom to the top, and drain valves installed so that the clear water could be decanted. A schedule of regularly decanting the tank was established. The solids content of sludge from the holding tank was increased to about seven percent.

After the problems with the clarifier sludge level controller described in Section VIA1 were corrected the sludge from the clarifier averaged about seven percent solids. Little improvement was then obtained by decanting the holding tank.

Considerable difficulty was encountered in reliably pumping the seven percent solids sludge from the hold tank to the incinerator. Plugging problem, similar to those described in Section VIA 1 under Clarifier Sludge Pump, occurred frequently. The steel pipe and ninety degree elbows were replaced with 2.54 cm. (1 inch) I.D. reinforced synthetic rubber hose, the gear pump was replaced with a variable speed drive progressing cavity pump complete with bridge breaker, and a procedure was established to flush the lines daily. Plugging was reduced to infrequent intervals.

Samples of Hydrasieve solids taken over a nine month period of steady state operation, July 1973 - March 1974, averaged seventeen percent total solids. The solids composition was about 72% volatile compared to about forty-eight volatile solids in the sludge. The clarifier sludge averaged 6.8% total solids and the hold tank sludge 6.7%. These data are contained in Table 18. The analyses of intensive evaluation samples performed by an independent laboratory contained in Table 19 are in agreement.

The wet weight of Hydrasieve solids collected during one week in April, 1973, averaged 18.2 kilograms (40.2 lbs). Average occupancy was eighty-two houses and average sewage flow was 102.66 cu. meters (27,122 gal.) per day. These data are contained in Table 20.

2. Dewatering

The original concept of solids dewatering and disposal, as described in detail in Section IV-B-1, Continuously Regenerating Filter, was for the Hydrasieve solids to drop from the screen through a hopper to a belt conveyer which fed a sand filter developed and built by the Procedyne Corporation. These solids together with sludge from the holding tank were to be mixed with hot recycle sand from the incinerator, dewatered in the filter to about eighteen percent solids via evaporation, vacuum filtration, and compression, then fed with a screw conveyor into the fluid bed of the incinerator.

Several problems developed:

a. The belt conveyor required frequent cleaning. Solids would accumulate on the drive and guide rolls resulting in poor tracking. Solids frequently would stick to the belt.

b. The overflow rate of sand from the incinerator to the filter was erratic, thus the mixing of hot sand, Hydrasieve solids, and sludge was not uniform. It was necessary to get good uniform mixing to cool the sand and evaporate part of the water content of the solids.

c. The filter operation was unsatisfactory. The sand-sludge mixture did not dewater and move through the filter as designed. The layer along the wedge wire screen would dewater and stay in place, blinding the screen, then dewatering would cease. Sometimes, when the sludge concentration was high, the sand-sludge mixture would form balls.

d. The feed frequently carbonized on the screw before being discharged to the fluidized bed thus plugging the screw.

It became apparent that further development work was required. In addition, as the Clarifier operation was improved and the sludge consistency approached seven percent solids, the economics of operating the sand filter became marginal due to the cost of power to run the filter and vacuum pump, and the cost of oil required to reheat the recycle sand from ambiant to 816°C (1500°F). The sand filter was removed.

TABLE 18

FREEHOLD SLUDGE CHARACTERISTICS

DATE 1973-74	Hydrasieve Solids		Clarifier Sludge		Hold Tank Sludge	
	% Total Solids	% Volatile Solids	% Total Solids	% Volatile Solids	% Total Solids	% Volatile Solids
July 31	14.2	71.5	5.5	43.4	6.8	46.8
Aug 6	11.3	92.6	5.9	48.7	5.5	41.5
Aug 16	12.9	77.9	8.3	48.6	6.1	52.6
Aug 21	16.7	68.5	7.5	50.3	9.8	58.8
Aug 26	12.6	67.7	10.6	73.0	4.0	52.1
Sept 5	16.4	85.6	5.8	53.6	-	-
Sept 10	27.7	40.3	8.4	46.5	9.8	46.3
Sept 20	23.4	49.4	9.1	49.4	-	-
Sept 25	16.2	86.5	7.0	44.1	5.2	53.9
Sept 30	16.1	71.0	9.1	53.1	8.5	53.0
Oct 10	16.4	67.6	9.0	40.7	6.8	44.7
Oct 15	13.4	81.6	3.6	43.1	8.6	41.7
Oct 25	31.9	34.9	8.5	46.2	10.2	41.0
Oct 30	23.6	59.6	2.9	50.7	3.4	47.7

TABLE 18 (Cont'd)

DATE 1973-74	Hydrasieve Solids % Total Solids	Hydrasieve Solids % Volatile Solids	Clarifier Sludge % Total Solids	Clarifier Sludge % Volatile Solids	Hold Tank Sludge % Total Solids	Hold Tank Sludge % Volatile Solids
Nov 8	13.2	84.0	7.5	44.0	7.0	44.7
Nov 14	14.3	90.0	4.8	41.5	5.8	45.0
Nov 19	11.9	70.8	-	-	-	-
Nov 27	14.0	71.2	6.3	49.1	5.6	48.5
Dec 5	15.7	80.5	6.3	45.5	7.8	45.4
Dec 14	17.4	94.1	4.7	48.0	4.4	48.6
Dec 21	25.2	44.0	9.5	41.0	8.3	49.6
Dec 26	11.2	82.5	5.5	48.1	9.4	46.9
Jan 21	16.2	86.7	6.2	56.2	6.1	56.2

FREEHOLD TREATMENT PLANT

TABLE 19

INTENSIVE EVALUATION - SLUDGE SAMPLES

DATE 1973	Hydrasieve Solids		Clarifier Sludge		Hold Tank Sludge	
	% Total Solids	% Volatile Solids	% Total Solids	% Volatile Solids	% Total Solids	% Volatile Solids
Sept 27	14.0	93.3	6.6	50.9	9.0	50.1
Oct 23	18.2	84.4	4.1	55.4	4.9	52.6
Nov 12	21.3	86.4	6.6	52.0	7.6	53.6
Nov 15	19.0	90.9	8.9	47.9	7.0	51.3
Dec 10	13.1	87.8	7.0	52.3	4.4	64.8
Dec 17	26.6	72.4	7.1	56.5	6.2	55.6
AVERAGE	18.7	85.9	6.7	52.5	6.5	54.7

FREEHOLD TREATMENT PLANT

TABLE 20

HYDRASIEVE SOLIDS COLLECTED

Date 24 Hour Period Ending		Solids Collected Wet Wt. lbs.	Sewage Flow Gallons
0830	4/12/73	45	24,315
0830	4/13/73	40	29,807
0900	4/14/73	24	28,611
0800	4/15/73	46	26,857
0830	4/17/73	46	26,019
AVERAGE		40.2	27,122

FREEHOLD TREATMENT PLANT

A vacuum coil filter was installed and tested for sludge dewatering. This unit was able to produce a filter cake of twelve to twenty percent solids with a sludge feed of five to ten percent solids. Suspended solids in the filtrate ranged from 6,000 to 20,000 mg/l. The filter required considerable operator attention and could not be left operating unattended for long periods of time. It was removed.

When the sand filter was removed it became necessary to use a different technique to feed sludge to the incinerator. A spare nozzle about 30.5 cm (12 inches) above the air distribution plate was fitted with a plug valve and a packing gland. A sludge feed gun consisting of a length of 1.25 cm diameter (0.5 inch) type RA 330 stainless steel pipe with a full bore ball valve on the outer end was inserted through the packing gland, plug valve and nozzle, extending about 10.16 cm (4 inches) into the sand bed. The sludge line with an air purge was connected to this gun. The small gun would plug once or twice a day. This gun was replaced with one 2.54 cm (1 inch) in diameter and an air purge rotometer was installed so the purge could be regulated at about 0.028 cubic meters per minute (1 SCFM). Gun plugging was no longer a problem.

3. Incineration

The "Fluidhearth" fluid bed incinerator, instrumentation, and related auxiliary equipment including that necessary for carbon regeneration was designed, fabricated, and installed under the supervision of the Procedyne Corporation. It is described in detail in Section IV-B-2.

Installation was completed in November 1972. The refractory was dried, thirteen hundred pounds of sand was charged to the unit and it was heated to 816°C (1500°F). The procedure developed to dry and cure the refractory whenever major alterations or repairs were made consisted of heating to about 71°C (160°F) with a propane torch, inserted through the recycle sand nozzle. After holding at 71°C for eight hours the temperature was increased to 110°C (230°F) at a rate of about 11°C (20°F) per hour, and held at 110°F for eight hours. The temperature was then increased to 316°C (600°F) at a rate of 28°C (50°F) per hour and held for four hours after which the regular start-up procedure was followed.

The original start-up procedure consisted of charging the incinerator with 590 kg (1300 lbs.) of 20-40 mesh flintshot sand, igniting the oil fired start-up burner in the plenum and heating to a bed temperature of 621°C (1150°F). The sand bed at this point was fluidized and the safety temperature interlocks allowed fuel oil to be injected directly into the bed via the gun burner. The bed temperature was increased to 816°C (1500°F), the start-up burner was shut down, and the controls set for automatic operation.

Shortly after start-up the differential pressure readings across the fluid bed and the air distribution plate indicated an absence of sand. The unit was cooled and examined. The ceramic plate had cracked, several ceramic tuyeres were broken, and the sand had drained into the plenum. The tuyeres were replaced with ones made of high temperature alloy (RA330) and a new ceramic plate was cast on top of the old one. The plate was cured, the incinerator was loaded with sand, and, following the regular procedure heated to operating temperature. Again, it was found that there was considerable leakage of sand to the plenum. In addition, several hot spots developed on the shell. Upon cooling the unit it was found that the packing around the plate had blown out. The plate was repaired by packing around the edge with "Fiberfrax" then building a refractory brick ledge to hold the packing in place. The hot spots were repaired by removing the refractory and deteriorated insulating brick in those areas and replacing with new materials. The unit was reheated and put on automatic control in late January, 1973.

During the spring and early summer of 1973 the incinerator was operated intermittently burning sludge first from the sand filter then directly from the sludge holding tank. The major operating problem was to maintain continuity of operations, especially during the sand filter operating period. The scrubber water circulating pump corroded and failed. It was replaced with a stainless pump. Oil tended to carbonize in the oil feed gun at low oil flow rates. An air purge similar to that on the sludge gun was installed. The fluidizing air blower failed and was replaced. The fuel oil pressure regulator failed and was replaced.

As the sewage flow to the plant and thus the amount of sludge collected increased, it became desireable for the incinerator to operate burning sludge automatically at night and at other times when no operator was in attendance. The sludge feed pump motor circuit was connected via relays to a high-low temperature controller which allowed the pump to run only when the fluid bed temperature was in the desired range of about 760^0 to 843^0C ($1400-1550^0$F). This worked very well.

During this period the carbon regeneration equipment was tested and made ready for operation. The incinerator carbon overflow valve was relocated to a better position. The carbon feed screw was redesigned to provide a new cooling jacket and stuffing box and a drain line was installed on the feed screw drain section. The after burner and controls were tested. The system was then put in stand-by because the carbon was not yet exausted and did not require regeneration. Later in the year it became apparent that the life of the carbon was much longer than originally believed. The incinerator was modified which precluded carbon regeneration.

On July 13, 1973 when restarting the unit after a power failure, a hot spot developed. An inspection revealed that the plate had cracked again. A 2.54 cm (1 inch) high temperature alloy type RA330 plate and tuyeres were installed. To insure fluidization in case of seal leakage a 3.18 cm diameter (1-1/4 inches) type RA330 air sparge pipe was inserted into the center of the bed from the top of the unit and valved to the fluidizing air blower. This system operated satisfactorily.

On August 31 the unit was examined by a licensed professional engineer from an independent laboratory while it was burning sludge at about design rate. Stack gas samples were taken as described in the New Jersey and Federal air pollution regulations. The results certified to the state show a particulate content of 0.011 grains/ft^3 and an opacity reading of zero.

During September the unit operated continuously and automatically to burn sludge, unattended at nights and checked only briefly on weekends. It shut down automatically on September 29, reason unknown, and was restarted September 30. The unit burned sludge faster than it was produced, thus when the holding tank level was low the sludge feed pump was stopped and the incinerator idled until the sludge inventory increased. During the month approximately 16,805 liters (4,440 gals.) of sludge at an average 7.8% solids was burned in 415 hours, a rate of 40.5 l/hr (10.7 gal/hr). Fuel oil consumption was 10.2 l/hr (2.7 gals/hr). Total ash removed from the dry cyclone was 360.6 Kg (795 lbs.) Fuel consumption averaged 730 gallons per ton of dry solids. At a fuel price of 18.4 c/gal the fuel cost per ton of dry solids was $130. These data are contained in Table 21.

The incinerator operated continuously until October 15 burning sludge as required. Air leakage around the plate seal became excessive and the unit was shut down. It was decided to relocate the start-up burner from the plenum to the top of the incinerator, remove the plate and in its place install an inverted concical shaped stainless distribution plate can. The tuyeres were on the top of the can and fluidizing air was piped from the blower through the plenum wall to the low point of the can. Sand was then added to the incinerator. The diameter of the distributor plate can was about 12.7 cm (5 in.) less than the inside diameter of the refractory thus the can was surrounded on all sides by sand and was free to expand and contract as the temperature increased and decreased. In addition there were no plate seals to erode. The unit was heated by down firing using the start-up burner on the top of the incinerator. The gun burner interlocks were changed so that the bed oil gun could be started when the temperature reached 566°C (1,050°F).

TABLE 21

INCINERATOR OPERATING DATA

SEPTEMBER 1975

Hours of sludge burning		=	415
Volume of sludge burned	gals.	=	4,440
Weight of sludge burned	lbs.	=	37,740
Sludge analyses	% Solids	=	7.8
Weight of dry solids burned	lbs.	=	2,944
Fuel oil used	gals.	=	1,120
Gallons fuel per dry ton solids		=	761
Cost of fuel	¢/gal.	=	18.4
Fuel cost per day ton solids	$/ton	=	130

FREEHOLD TREATMENT PLANT

The incinerator was restarted in February 1974. The new air distributor plate can concept worked well. Bed fluidization appeared good, however hot spots developed at points on the shell above the oil and sludge feed gun nozzle. The cast refractory at these nozzles was found to be cracked and the insulation eroded. This was repaired by drilling holes and pumping in castable insulation which sealed the cracks and filled the eroded voids, however upon subsequent operation, hot spots developed at other places on the shell. It was concluded that to make the unit operable the original brick-type insulation should be removed and the area between the refractory brick and the shell be filled by pouring castable insulation.

By the spring of 1974 the price of fuel oil and power were twice that in September 1973 and economics of operating such small units dictated hauling the sludge to landfill as long as sites were available and regulations permitted. It was finally decided to remove the incinerator and use the space for other purposes.

C. PLANT MODIFICATIONS

During the 2 year period since the spring of 1974 an attempt has been made to achieve reduced operating costs, lower attendant requirements, achieve greater consistency of results and measure the local community acceptance and reaction to the facility. Listed plant modifications are as follows:

1. Installed automatic hot water spray system for the Bauer Screen with timing devices controlling frequency and duration of flushing to eliminate grease and solid buildup on screens.

2. Installed a retention loop consisting of water storage tanks in series after the surge tank and prior to the clarifier for chemical mixing.

3. Added positive displacement pumps to the pump station eliminating the problems of the prior type which required frequent operator attention and caused recycling in the plant during periods of ailure.

4. Eliminated the magnetic filter with its head loss and power needs.

5. Removed the incinerator.

6. Installed a water pump to permit more efficient backwash with effluent rather than fresh water.

7. Replaced the carbon column supports with stainless steel and modified piping to simplify maintenance.

8. Added 2" decanting line from Bauer screen hopper.

9. Installed a 4,860-gallon ferric chloride storage tank to permit bulk delivery and allow automated feed and mixing.

10. Installed pH meter.

The above modification program has significantly reduced operating costs identified in the financial section.

SECTION VII
FINANCIAL CONSIDERATIONS

INTRODUCTION

Treatment plants have two major elements of cost (capital and operating costs) which are of concern to both the initial and ultimate owner. Funding, initially, is often the burden of a developer, but ultimately, all costs must be borne by the homeowner. The major problem is to balance the capital and operating expenses to achieve the best economic solution for the project. In today's society with changing criteria and more stringent quality requirements for effluent discharge, together with frequent moratoriums on new connections to the sewage system, a simple low labor intensive treatment plant producing a high quality effluent is very desirable. By low labor cost we are reflecting the desire to utilize part-time operators for four hours or less per day on a six-day basis with alarm systems and dualization of key elements to avoid breakdown problems. Such a plant can be justified to permit residential growth if the capital cost per dwelling unit and subsequent operating costs fall within the following ranges:

1. capital cost: $500-1,900 per dwelling unit
2. operating cost: $100-200 per dwelling unit per year

The above costs must be weighed against all alternates in the following situations:

1. Cost of pump station and force main to a trunk sewer.
2. Will interim facilities meet all quality needs and if so can the effluent meet stream discharge permitting requirements or will ground water recharge have a more desirable environmental impact.
3. Will the facility permit use of land that is currently unusable and is the enhanced land value greater than the total plant cost.

The system being constructed may be for an interim facility which will be ultimately abandoned, or it may be one which will have an extended life, requiring modifications to provide for future growth and for changing criteria. Such a plant may initiaially be owned and operated by a developer and may ultimately be deeded to or purchased by a government agency. These factors should be considered in the design of the proposed plant as its future fate will effect choice of process and equipment and lead to design considerations which will be directly related to the ultimate costs to the customer.

In this project of extremely limited size a major attempt was made to arrive at lowest potential operating costs while at the same time recognizing that capital costs, although important, were not the major variable consideration.

SIGNIFICANT COST DATA

During the research phase of this project extensive cost records were maintained which are not significant in the long-range operational mode because of the high cost of sampling, most of which would not be required for routine operation, and because the continuous modification occurred in this experimental plant.

Subsequently, the second series of costs were developed which indicated serious underestimating of operating cost in the original designs due to the following:

1. Plant design produced excessive chemical need requirements;

2. Major increase in utilities cost because of fuel adjustment and rate increases to the power company;

3 Increased cost of chemicals.

4. Significantly increased labor costs.

The estimated costs of Table 22 (top set of numbers) were prior to construction and the actual costs in Table 22 (middle set of numbers) were during the evaluation phase. Subsequently, the following major modifications were made in the plant to improve operations, reduce labor, and permit lower purchase prices for the necessary chemicals:

1. Positive head pumps were installed at the lift station to reduce recycling and thus eliminate a major maintenance problem.

2. Underground bulk storage of ferric chloride was provided reducing its purchase price and handling costs to less than half of those previously incurred.

TABLE 22

OPERATING COSTS
FREEHOLD TREATMENT PLANT

ESTIMATED COST(1)
EPA GRANT APPLICATION 1970

ITEM	COST $/1,000 GAL.
Direct Labor (8 hrs./week)	$ .092
Materials (includes chemicals)	.134
Electric	.208
Gas	.038
Maintenance	.046
	$0.518

ACTUAL COST(2)
STEADY STATE OPERATING PERIOD
JULY 1973-MARCH 1974

ITEM	COST $/1,000 GAL.
Labor (6 hrs./day, 7 days/week)	$ .793
Chemicals	.400
Electric	.266
Fuel Oil	.150
Maintenance	.066
Sludge Hauling	.350
Telephone	.050
	$2.075

PRESENT COST(2)
1976 OPERATION

ITEM	COST $/1,000 GAL.
Labor (4 hrs./day, 7 days/week)	$ .57
Chemicals	.27
Electric	.32
Fuel Oil	.08
Maintenance & Supplies	.20
Sludge Hauling	.39
Telephone (Ans. Service)	.02
	$1.85

(1) Cost based on 50,000 GPD

(2) Cost based on 28,000 GPD

3. Major modifications were made in the flow pattern permitting increased mixing and detention time prior to the clarification thus reducing the quantity of chemicals required per 1,000 gallons of flow treated.

4. pH control of the process permitted reduced (required) dosage of ferric chloride.

5. The magnetic filter was eliminated.

6. Adjustments were made to improve the backwash capability in the carbon column.

The modifications cited above resulted in reducing the operating costs to the levels recorded in Table 22 for 1976 operations (lowest set of numbers).

IDEAS ON FURTHER COST REDUCTION

Based on the experience with this plant, further cost reductions are possible if initial designs incorporate the following concepts:

1. A new configuration of the plant can reduce labor needs and energy costs.

2. Proper design with equipment selected for long life and minimum maintenance will reduce the operating cost.

3. Increased sludge tank holding capacity will decrease the frequency of trucking sludge and also reduce costs.

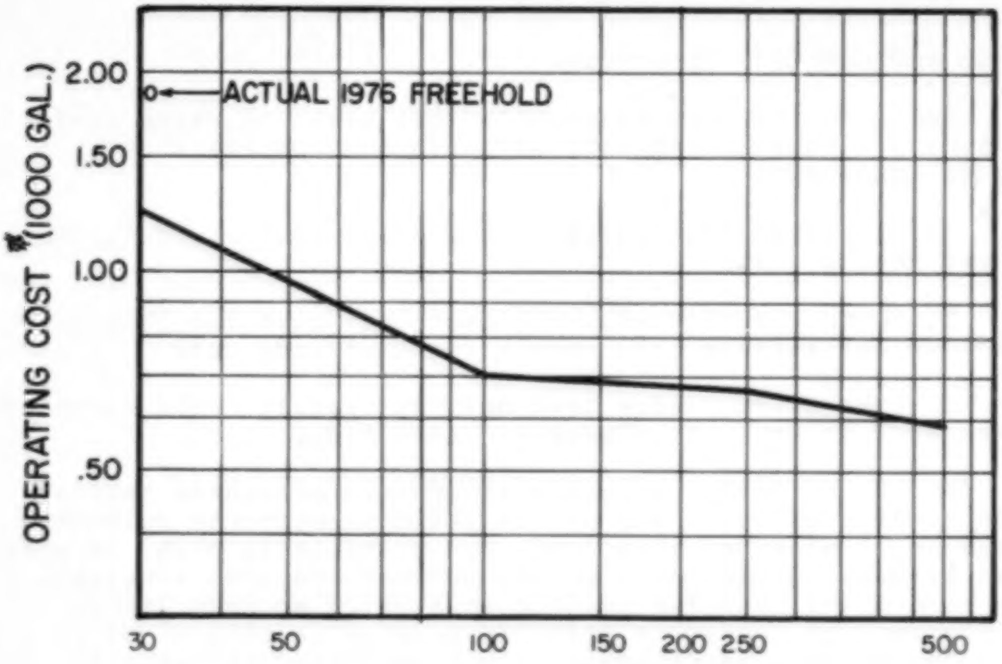
4. Using ferric chloride to achieve phosphate removal in high quality effluent may not be the most economic selection of coagulant. At other locations, other chemicals might be used such as primary polymer or alum and polymer and thus simultaneously reduce the need for caustic soda which we have had to use to correct pH. (Ferric chloride as used in this plant reduced the pH to a point where it was necessary to raise it with sodium hydroxide in order to provide the optimum pH for coagulation).

5. The substitution of alternate coagulants for the ferric chloride would eliminate the need for PVC piping and epoxy coated tank linings and thus reduce total equipment and mixing requirements and costs.

Based on the above considerations a projection of operating costs independent of capital costs and independent of inflation costs has been made for different size plants as shown on Figure 43 and Table 23.

FIGURE 43

ESTIMATED
OPERATING COST vs. PLANT SIZE



PLANT SIZE GPD x 10^3

FREEHOLD TREATMENT PLANT

TABLE 23
PROJECTED OPERATING COSTS

$/1,000 GALLONS

ITEM	PLANT SIZE 55,000 GPD	100,000 GPD	250,000 GPD	500,000 GPD
LABOR	$ .28	$ .18	$ .16	$ .14
ELECTRIC	.18	.14	.14	.12
FUEL OIL	.04	.03	.03	.02
REPAIR AND MAINTENANCE	.16	.12	.12	.10
CHEMICALS				
FERRIC CHLORIDE	.10	.08	.08	.07
SODIUM HYDROXIDE	.13	.12	.10	.10
CHLORINE	.02	.02	.02	.02
POLYMER	.02	.02	.02	.02
TOTAL OPERATING COST/1,000 GALLONS	$ 0.93	$ 0.71	$ 0.67	$ 0.59

FREEHOLD TREATMENT PLANT

APPENDIX A

PROCEDURES FOR ANALYZING FREEHOLD SAMPLES

Freehold samples which were analyzed at the Marshallton lab were performed in accordance with the references and procedures listed below. Where a test is referenced to Standard Methods or EPA, it was done as directed in the reference. Any modifications of a standard are listed. Those tests which are performed by some technique other than Standard are listed with backgrounds and explanations for their use.

Suspended Solids:	Standard Methods for the Examination of Water and Wastewater, 13th Edition (1971) Method 224 C, p. 537.
Coliform:	Standard Methods, Method 407 A & D p. 664.
Chemical Oxygen DemandL	Standard Methods, Method 220, p. 495.
Total Solids:	Standard Methods, Method 224 A, p. 535.
Volatile Solids:	Standard Methods, Method 224 B, p. 536.
Alkalinity:	Standard Methods, Method 102, p. 52, modified only by using methyl purple instead of methyl orange as indicator.
Chloride:	Standard Methods, Method 112 A, p. 96.
Biochemical Oxygen Demand:	Standard Methods, Method 219, p. 489, modified for Weston and Stack Dissolved Oxygen Probe as approved by Methods for Chemical Analysis of Water and Wastes, EPA (1971), p. 60.
Phosphorus:	Methods for Chemical Analysis of Water and Wastes, EPA(1971) pp. 235-245.

Nitrate and Nitrite: Water Analysis Handbook, Hach Chemical Co. (1973), modified from Standard Methods, Method 213 B, p. 458.

The Hach procedure is like Standard Methods in that it is a photometric color adsorbtion of a sample in which the nitrate has been reduced by calmium. However, the Hach test is much faster and more convenient because it employs reagent powder pillows instead of a reduction column.

A Nitra Ver IV Powder Pillow is added to 25 ml of sample and shaken for one minute. If nitrate or nitrite are present, a pink color will develop which is measured against a blank in a Fisher Electrophotometer II using a greem filter with a wavelength of 525 m/l. The number from the Photometer scale is then applied to a calibration curve from which the Photometer scale is then applied to a calibration curve from which Nitrate/ Nitrite - Nitrogen is read in mg/l.

Total Oxygen Demand: Instruction Manual Model 225, Total Oxygen Demand Analyzer, Ionics, Inc. (1970).

The Ionics Model 225 TOD Analyzer is an automatic accurate instrument capable of graphically recording the oxygen demand of dissolved and suspended oxidizable constituents in an aqueous sample. Its advantages over other common methods are that it is rapid, automatic, more precise, and has a chemical reaction efficiency at or approaching theoretical. The instrument can be set to perform within oxygen demand limits of 0-50 ppm for the low range, extending to 1000 ppm for high range applications.

A 20 microliter sample drop is automatically injected into a furnace tube containing platinum catalyst at 900°C. A nitrogen stream containing a fixed, known amount of oxygen continuously flows through the furnace tube and supplies the oxygen necessary for combustion. The vaporized, oxidized sample along with the oxygen depleted nitrogen stream, then passes through a fuel cell containing platinum and lead electrodes in a medium of 20% KOH solution. The fuel cell reacts to the loss of oxygen in the nitrogen stream and converts this imbalance to electrical impulses which are translated to a chart read-out from which oxygen demand is read in mg/l TOD.

Total Organic Carbon: Analytical Chemistry, Vol. 37, No. 6, May 1965, p. 769; Vol. 39, No. 4, April 1967, p. 503.

The same sample which is oxidized in the TOD analyzer passes through a MSA Lira Infrared Analyzer Model 300 wherein the TOC of the sample is measured and recorded. Prior to sampling by the TOD analyzer, the sample is adjusted to pH<3 and then sparged with nitrogen to remove inorganic carbon compounds, thus leaving only the organic carbon compounds to be oxidized to CO_2 and measured by the infrared optical bench and graphed as mg/l TOC.

Ammonia: Refer to letter to J. D. Beach from R. A. Conte, dated November 19, 1974, RE: Analysis for Ammonia Nitrogen.

Add about 50-80 ml of sample into a 100 ml beaker containing a stirring bar. To this, add 1 ml of 10 M NaOH for each 100 ml of sample. This addition raises the pH of the sample above 11 so that all of the ammonium ion is converted to ammonia which can be detected by the probe. The probe is then inserted into the sample solution which is agitated by a magnetic stirring bar. The digital millivolt read-out is then applied to a calibration curve which yields mg/l ammonia. Multiply mg/l ammonia by 0.82 to get mg/l ammonia nitrogen. Rinse probe well between samples.

To prepare a calibration curve, use serial dilutions of ammonia chloride for samples containing 100, 10, 1, and 0.1 mg/l ammonia. Using the 100 ppm ammonia sample and above procedure, set the meter at -75 mv. Then run the remaining standards and record the mv for each. Plotting these data points on 4 cycle, semi-log paper should give a straight line.

Total Kjeldahl Nitrogen: Refer to letter to J. D. Beach from R. A. Conte, dated November 19, 1974, RE: Analysis for Ammonia Nitrogen.

This Kjeldahl test converts organic nitrogen to an ammonium ion state which can then be readily converted to ammonia and read by the Orion Ammonia Probe.

Using a micro Kjeldahl apparatus, a 50 ml sample is digested in the presence of a digestion reagent (see Standard Methods, 13th Edition, p. 245, 3.a.). The digestion is carried out until SO_3 fumes appear and then subside. After cooling, the digestion flask is rinsed with distilled water into a 100 ml volumetric flask. After several rinsings, the volume is made up to 100 ml. From this, a 20 ml aliquot is put into a 150 ml beaker to which is added 0.5 ml of a Saturated KI solution and 78 ml distilled water. The beaker with a stirring bar is then placed on a magnetic stirring motor and the ammonia probe is inserted into the solution. Add 1.5 ml of the 10 M NaOH and read the millivoltage; then add 10 ml of a 76.4 mg/l. NH_4Cl solution and again read the millivoltage on the meter.

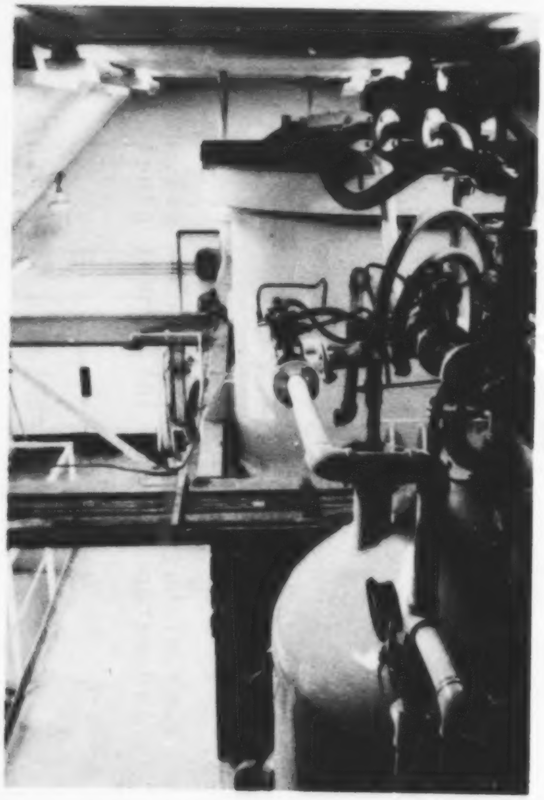
This potential difference is applied to a known addition chart supplied with the ammonia probe. The chart reading is then multiplied by 200 and this is reported as mg/l Total Kjeldahl Nitrogen.

APPENDIX B

PHOTOGRAPHS
FREEHOLD TREATMENT PLANT
DECEMBER 1975

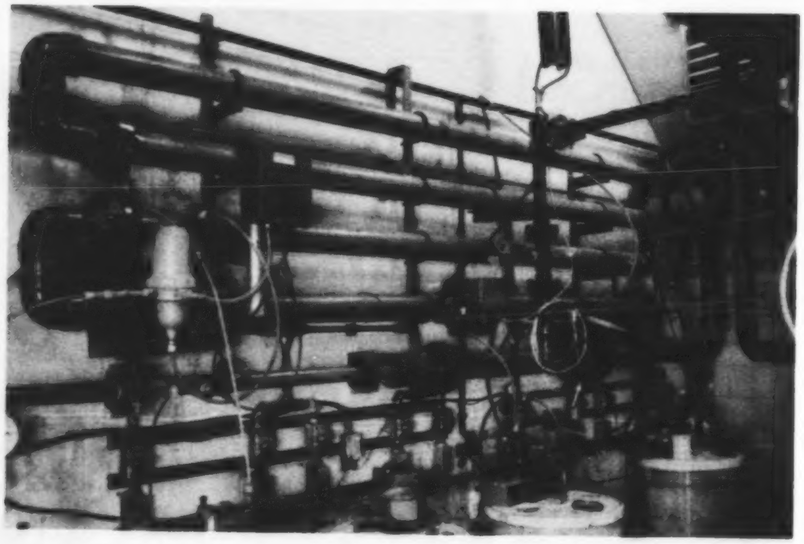
VIEW OF PLANT AND MANASAQUAN RIVER TREE LINE

VIEW OF PLANT EFFLUENT LINE INTO MANASAQUAN RIVER

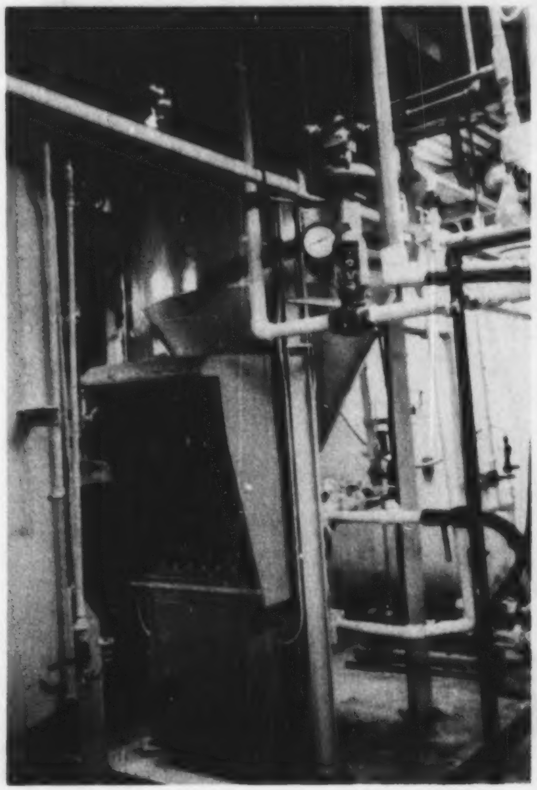


← View of main floor from carbon column to clarifier to chemical mixing loop against wall on the far end of building. Entrance door is to the left at the far end of building.

View of chemical mixing loop with ferric chloride, caustic soda and polymer tanks. ↓



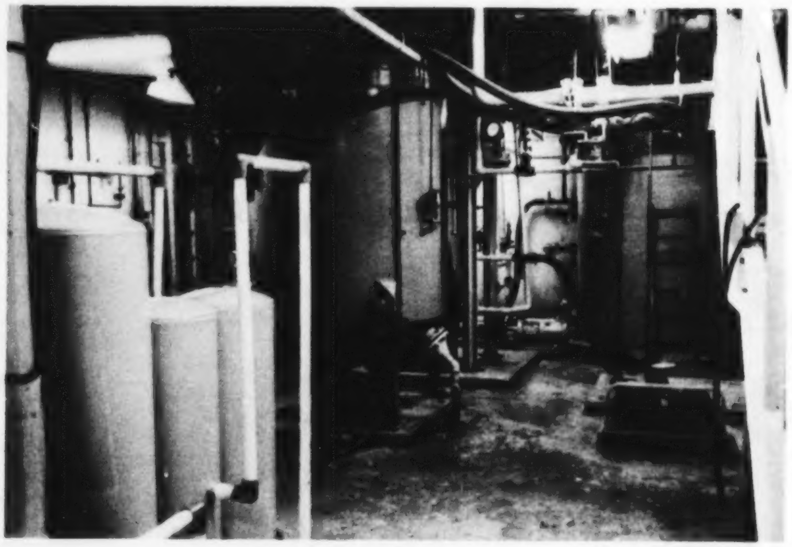
FREEHOLD TREATMENT PLANT



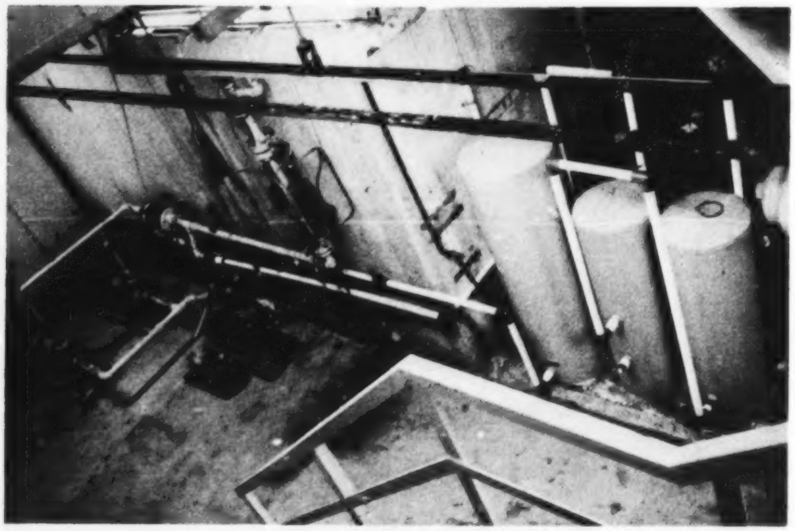
View of main floor instrument panel;
carbon column and door leading to
office and laboratory

View of basement floor carbon column
and carbon column instrument panel

FREEHOLD TREATMENT PLANT



View of basement floor with retention loop for chemical mixing to immediate left, sludge holding tank far left and carbon column feed tank far right



View of basement floor with retention loop for chemical mixing to the right and primary feed pumps to the far left

FREEHOLD TREATMENT PLANT

APPENDIX C

CONVERSION TABLE(1)

TO CONVERT FROM	TO	MULTIPLY BY
foot	metre(m)	3.048000 E-01
ft^2	metre^2(m^2)	9.290304 E-02
ft^3	metre^3(m^3)	2.831685 E-02
gallon(U.S. liquid)	metre^3(m^3)	3.785412 E-03
gal(U.S. liquid)/min.	metre^3/sec(m^3/s)	6.309020 E-05
grain(1/7000 lb avoirdupois)	kilogram (Kg)	6.479891 E-05
horsepower(electric)	watt (W)	7.460000 E+02
inch	metre(m)	2.540000 E-02
in^2	metre^2(m^2)	6.451600 E-04
in^3	metre^3(m^3)	1.638706 E-05
pound(lb avoirdupois)	kilogram(Kg)	4.535924 E-01
ton(short,2000 lb)	kilogram(kg)	9.071847 E+02
yard	metre(m)	9.144000 E-01
yd^2	metre^2(m^2)	8.361274 E-01
yd^3	metre^3(m^3)	7.645549 E-01
degree Fahrenheit	degree Celsius	$t^{o}{}_{c}=(t^{o}{}_{f}-32)/1.8$

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(1) As per Standard for Metric Practice E380-76
American Society for Testing and Materials